

CHEMICAL ENGINEERING AND THERMODYNAMICS

APPLIED TO

THE CEMENT ROTARY KILN

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CHEMICAL ENGINEERING AND THERMODYNAMICS

APPLIED TO THE CEMENT ROTARY KILN

*A TREATISE FOR KILN-DESIGNERS,
ENGINEERS, WORKS CHEMISTS, AND
WORKS MANAGERS, EMBRACING RECENT
RESEARCH WORK, WASTE-HEAT BOILERS,
DUST KILNS, AND NEW DEVELOPMENTS*

CHECKED

BY

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INTRODUCTION

IN this work the complete theory of the rotary cement kiln has been worked out, based in the main upon thermodynamical calculation, in which the recent advances in physical chemistry have been taken into account.

As a result we are enabled to calculate the correct clinker output from a rotary kiln of any given diameter—ranging from 1 to 20 ft. if need be—the theoretical outputs arrived at by thermodynamics being checked against actual practice. Also the sizes of the clinkering and decarbonating zones necessary for a given output are determined, these sizes being also worked out by thermodynamics and checked against actual yields of kilns. Also the correct dimensions of rotary kilns are worked out, including rules for finding their lengths, diameters, and sizes of the various zones of both wet and dry process kilns. Also the correct air supply is worked out, and the entropy of cement formation.

The value of such data is obvious because on the one hand it enables the works manager—knowing the dimensions of his kiln—to know whether or not the kiln is being worked at maximum capacity and economy. It also enables the engineers who construct kilns to design them with certainty and guarantee output of any size.

Most of this work is new, being based upon researches carried out by the author while Director of Research of the British Portland Cement Research Association. All the work has been carefully checked against actual practical results, and will be found reliable.

It may at first sight appear surprising that, since the rotary kiln has been in almost universal use since 1906, this work was not done previously. There were, however, many technical difficulties in the way. In order to accumulate experimental data, the British Portland Cement Research Association carried out no less than twenty-nine kiln tests on every variety of kiln, at first under the able direction of Mr William Gilbert, and later under that of the author. The masses of experimental data arising out of the tests had then to be collated; tables of specific heats and other physical properties of the gases and raw materials which occur in the process of manufacture had to be calculated, and in some cases experimentally determined at high temperatures.

Exact allowances had to be made for radiation and convection effects of complicated nature. All this preliminary work represented immense labour and great cost.

A few lacunæ still exist in our scientific data, but enough scientific and engineering data have been accumulated to enable important deductions and suggestions to be made.

§ 3. Perhaps the most striking result brought to light in the following pages is the degree of inefficiency of the rotary kiln.

For many years practical men have been aware that the rotary kiln is inefficient, but how and where inefficient could not be exactly known until the details worked out below were ascertained.

On the basis of the calculations presented, the efficiency of rotary kilns is barely 19 per cent.*

* It is shown below that an average rotary kiln consuming, say, 33 tons of standard coal per 100 tons of clinker should, if theoretically perfect, consume only 6.35 tons. Defining efficiency

$$E = \frac{\text{Possible fuel consumption}}{\text{Actual fuel consumption}} \times 100.$$

$$\text{we obtain} \quad E = \frac{6.35 \times 100}{33} = 19.2 \text{ per cent.}$$

§ 4 It is obvious, therefore, that there is present in the cement industry a wider scope for improvement both of kiln design and for research for improved methods of production than in almost any other manufacture

One important practical result which is brought to light below is the fact that a rotary kiln fed with a slurry containing 40 per cent of water should work with a fuel consumption of about 21 tons of standard coal per 100 tons clinker produced

By reducing the slurry moisture below 40 per cent, improvements beyond this could be registered

§ 5 Another result which is emphasised by the inquiry below is the supreme importance, from the point of view of fuel economy, of attaining as high a flame temperature as possible in the clinkering zone. Such a high temperature cannot be attained without special precautions, chief among which is the preheating to as high a temperature as possible of the air entering the furnace

The efficient preheating of entering air is by no means easy, but, in view of recent advances in furnace design, does not present the same difficulty as formerly

§ 6 Perhaps, however, the most important result is the discovery of the pre dominant part played by radiation and convection—both internal and external—in causing fuel losses. These losses are very much more serious than has hitherto been imagined, and it is only by their elimination by suitable kiln design that striking reductions in fuel economy will result

§ 7 Possibly the main reason why this influence has not been sufficiently recognised in the past is the fact that writers on the rotary kiln have been accustomed to regard all their heat units as of equal value. Heat balances of the kiln, such as have been repeatedly drawn up during the last thirty years, do not indicate that the B Th U's available at different temperatures have very different values both chemical and financial. For example, only these B Th U's available *above* a temperature of 1481° F (805° C)—the temperature of decomposition of calcium carbonate in the kiln—can be directly utilised for forming clinker. Consequently, they are of great value to the cement manufacturers

The B Th U's available below those temperatures are, so to speak, merely waste products so far as clinker formation is concerned. They merely serve to preheat the entering slurry

In view of this fact the ordinary heat balances of the rotary kiln are somewhat misleading in that all the B Th U's in them are tabulated as if they were of equal value and of equal cost

It is as if in making up a financial statement of an account we took, say, 1,000 coins and regarded them as all of equal value. There may be gold, silver, or copper coins among these 1,000 yet this factor is neglected in making up the balance sheet, which simply tabulates the number of coins which was paid for each article irrespective of the value of the individual coins. This procedure would not meet with the approval of commercial men, yet it is precisely the *modus operandi* of the writers who have drawn up the thermal balance sheets of our kilns in the past

§ 8 For example in the ordinary heat balance of the rotary kiln the external radiation losses on the whole available B Th U's work out about 8 per cent

If, however, we work out the percentage of these radiation losses in respect to the valuable clinker forming B Th U's—the only ones which count—we find that nearly 15 per cent of the effective heat available for clinker formation thus uselessly disappears as external radiation. This alters entirely the complexion of the question of whether it is worth while stopping external radiation from the clinkering and decarbonating zones. We are, in fact, wasting nearly 15 per cent

of our fuel on account of external radiation from the hotter zones of the kiln, and therefore the greatest trouble should be taken to stop such losses.

Moreover, the loss of clinker output from radiation and convection *inside* the kiln is very much more serious than the loss due to external radiation. The importance of these internal losses does not seem to have been realised by practical men. The kiln is a tube with one end red hot and the other end comparatively cold. High-grade heat is radiated away from the red-hot decarbonating and clinker-ing zones directly into the colder parts of the kiln without doing any useful work. In addition to this, much high-grade heat is carried away bodily by the hot gases from the decarbonating zone, passing into the colder zones. Since every 918.6 B.Th.U.'s of high-grade heat (*i.e.*, heat available above 1,481° F. or 805° C.) thus carried away into the colder parts of the kiln represent a loss of 1 lb. of potential clinker, it is obvious that no great progress in kiln economy can result until the design is altered so as to reduce such losses.

For these reasons I have applied to cement formation what has long been done for steam by steam engineers—namely, the conception of entropy.

§ 9. In the following pages we have reduced to a matter of exact thermodynamical calculation most of the quantities connected with the rotary kiln.

Thus we have, for the first time, calculated exactly the exit temperatures which the gases issuing from a rotary cement kiln will possess when the coal consumption for 100 tons clinker is reduced to any given figure, and when it is fed with a slurry containing any given percentage of moisture. This, previously, it has been impossible to do.

As a necessary consequence we can now calculate beforehand the amount of steam producible from these exit gases under all possible conditions, ranging from a dry process kiln to a kiln fed with a slurry containing, say, 50 per cent. of moisture.

Of course the converse proposition has long been solved—that is, *knowing the exit temperature and the composition* of the gases, it has been possible to calculate the amount of steam that can be produced thereby—but it has not been possible up to date to calculate the exit temperature appertaining to any fuel consumption.

A chapter on the design of waste-heat boilers has been added, giving the amount of steam which can be raised therefrom under various conditions—a subject on which the writer has had much practical experience.

§ 10. The determination of exact scientific data is the first step towards modern technical advance. In Sir Charles Parson's words:

“The hopelessness of making progress in the absence of data and an intimate knowledge of a subject was deeply impressed on the minds of those who served on the Board of Invention and Research during the war, and who received at one period more than 2,000 inventions and suggestions per week, of which only a minute fraction of 1 per cent. were in any way helpful, while scarcely any were of practical value. The immense wastage of mental effort appeared to them deplorable.”

From this point of view it is hoped that the data brought together in this book will prove indispensable to kiln designers, engineers, and works chemists. Indeed, without such scientific data it is difficult to see how any really efficient kiln could be designed, as the engineers have never been presented with many of the essential facts.

§ 11. A few words on the origin of this work will prove interesting.

The main portion of this work is based upon research work carried out by the author while Director of Research of the British Portland Cement Research Association, between 1922 and 1925. It will be recollected that during the war, and under the compelling fear of the scientific ability of the Germans, the British cement manufacturers had been persuaded by the Government, by means of a handsome donation, to combine and subsidise scientific research on cement kilns,

and they founded the *British Portland Cement Research Association*. With the removal of the German danger, and in view of the enormous profits made by the British industry owing to their virtual monopoly and excellent raw materials all incentive to progress stopped and the Research came to a sudden and premature end in 1925 just when fundamental progress was beginning to be made in unravelling the intricate chemical and physical phenomena underlying the manufacture of cement. This disaster was largely the result of the tragic death of Mr H. D. K. Bamber which at a critical time removed the foremost advocate of scientific research and progress from the British cement industry.

With the dispersal of the scientific staff and facilities for research all hope of completing the numerous scientific researches vanished. The author, however, has privately and under great difficulties completed some of the research work relating to rotary kilns and this book contains some of the data he has been able to rescue from oblivion in the general wreck. It will be obvious from a perusal of this work that the science of Cement Manufacture is as yet but in its infancy.

Many cement manufacturers have mistaken purely routine work for research work. True research work on cement is costly and presents many difficulties. The product is formed at an immensely high temperature and consists of a mixture of refractory silicates in a state of what Dr Mellor calls 'arrested equilibrium'. The problems presented are of the highest order of scientific difficulty, involving Physical Chemistry, Physics, Mechanics, Thermodynamics, Engineering and Geology. Fundamental advances will never be achieved by employing a few half-trained hands driven under harassing and anxious factory conditions, pressed for immediate financial results in an inimical and sceptical atmosphere.

It has often been claimed that 'rule of thumb' methods have been more fruitful of results than purely scientific methods in the cement industry. The answer to this statement is that scientific efforts on a very difficult subject have never been carried out for a sufficient length of time nor under the free conditions requisite for achieving the astounding results attained in other branches of applied science such as synthetic dyes, synthetic fertilisers, electrical work and aeronautics. On the other hand 'rule of thumb' methods have been energetically pursued for over a hundred years in the cement industry by enormous numbers of workers. When we reflect that three or four university-trained scientists, working for the very few years between 1922 and 1925 for the British Portland Cement Research Association under most harassing and anxious conditions discovered the Laws of the Science of Fine Grinding (which have received world-wide acknowledgment and confirmation) and for the first time in history worked out the whole theory of the rotary kiln and the causes of its inefficiency we must conclude that given adequate encouragement a few years more of research would have seen the discovery of revolutionary improvements thus bringing the British cement industry into line with other modern industries.

It is sad to acknowledge that the prevailing spirit towards research in this industry in the twentieth century amidst unparalleled progress in all branches of applied science was such that in a time of great prosperity a sheltered, wealthy and powerful trade actually celebrated the centenary of the discovery of Portland cement by Aspin by destroying its Research Association and annihilating its research men although the expense of the research worked out at only about one fourteenth of a penny per ton of cement produced!

The thought lies near that possibly it is this short-sighted mental attitude towards new developments not only in the cement trade but in other industries as well that is partly responsible for the steady loss of British trade for so many years past.

Another objection which has often been urged against scientific research in the British cement industry is that it is so complicated that the cement manufacturer cannot understand it, not even comprehending the meaning of the language in which the results are expressed. The only method of removing

this difficulty lies in improved technical education. There are very few university-trained scientists in the cement trade, and it is elementary psychology that what the human mind does not understand, and what is strange to it, it dislikes and opposes.

There is no way of simplifying science, or of altering the complicated methods whereby Nature works her irresistible way, blind to all human needs, along to unseen ends. All technical men, and most manufacturers, feel inclined to sympathise with Louis XIV. when he decided that had he been consulted in the Beginning of Things he would have greatly improved the universe by providing it with a set of much simpler laws!

In any case it is no use blaming the poor scientist for the complications of the universe, as some cement manufacturers have done.

This work owes its publication to the invitation of Mr Nathan C. Rockwood to contribute a series of articles to "Rock Products" on the rotary kiln, and most of the material in this book has appeared in the pages of this American journal.

I have also to thank Mr Childe, the Editor of *Cement and Cement Manufacture*, for leave to reproduce my article on "The Entrophy of Cement Formation," which first appeared in that journal, and also for kindly allowing me to reproduce the illustration of the "Waste-heat Boiler," Fig. 13 of Chapter XXVII.

I desire also to thank Dr Mellor, F.R.S., for leave to reproduce part of my articles and illustrations on Flame Temperatures and Waste-heat Boilers, which first appeared in the *Transaction of the Ceramic Society*.

GEOFFREY MARTIN.

LONDON.

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CHAPTER I

MEASUREMENT OF THE COAL CONSUMPTION
OF CEMENT ROTARY KILNS BY MEANS OF
ANALYSIS OF THE EXIT GASES

In making Portland cement clinker, part of the carbon dioxide in the exit gases arises from the combustion of the coal and part from the decomposition of the carbonates in the material from which Portland cement is made.

It is possible by accurately analysing the exit gases to estimate the coal consumption per 100 tons of clinker made, without the necessity of weighing either the coal or the clinker. The carrying out of a kiln test, in which the coal and the clinker are actually weighed, is, on most works, a laborious operation. Consequently, methods which obviate this necessity are valuable, since they enable the manager to check the fuel consumption of the kiln and control the work of his burners.

Savings of fuel to the extent of hundreds of tons a year could be effected in certain works merely by a rigorous control of the exit gas analysis, followed by a computation of the fuel consumption per 100 tons clinker made.

This may be achieved by several methods:—

Method 1.—The following method was devised by the writer:—

(1) From the analysis of the coal used on the works, calculate the weight of CO_2 and N_2 evolved per 1 lb. of coal burnt.

(2) From the chemical composition of the dried slurry, calculate the weight of CO_2 evolved per 1 lb. of clinker made.

(3) By now assuming different fuel consumptions, calculate the volumetric composition of the exit gas, and therefrom calculate the volume of air-free nitrogen for different fuel consumptions.

Then the fuel consumption is a function of the volume of the CO_2 per 100 volumes of air-free nitrogen in the exit gas, and by means of a table or curve the fuel consumption per 100 tons of clinker made can be read off for any given $\text{CO}_2\text{-N}_2$ ratio.

Now, in working out this formula, there are a number of factors which cannot be readily estimated, and which, on the whole, tend to make the practically found volume of CO_2 greater than the theoretical value. Thus the clay contains a certain amount of organic matter (humus, peat, woody fibre, etc.) which evolves carbon dioxide when burnt. The ash in the coal adds on to the weight of the clinker. There is a certain amount of sulphur in the coal which is partially estimated with the CO_2 .

There is a small amount of free nitrogen in the coal. The net result of these factors is that when the actual fuel consumption (with weighed quantities of coal and clinker) is practically determined and checked against the theoretical volume of CO_2 as calculated, it is found that the theoretical volume of CO_2 must be increased usually by about 4.5 per cent. in order to yield results which agree accurately with those found on practical tests.

Hence, for a given works it is advisable to construct a table as indicated below, and then check against a single determination of the fuel consumption with

weighed quantities so as to ascertain the actual percentage correction to be applied to the theoretical volume of CO_2 as calculated. If, then, a second practical table be constructed in which this correction is applied, it will be found that the actual fuel consumption can then be read off from this table accurately minute by minute, with very little calculation or trouble, merely by making accurate analyses of the exit gases by means of an Orsat apparatus.

TABLE

Showing Fuel Consumption Corresponding to Exit Gas Analysis

| (1) Number of Tons Standard Coal Consumed per 100 Tons Clinker Made (1 Lb Coal Gives 12 600 B Th U's) | (2) Theoretical Volume of CO_2 per 100 Volumes N_2 in Air Free Exit Gas | (3) Corrected Volume of CO_2 per 100 Volumes N_2 in Air Free Exit Gas Factor = 1.027 |
|--|---|---|
| Tons | Volumes | Volumes |
| Infinite | 21.66 | 22.9 |
| 90.01 | 26.45 | 27.95 |
| 45.77 | 31.10 | 32.87 |
| 41.00 | 32.20 | 34.04 |
| 39.23 | 32.67 | 34.53 |
| 35.30 | 33.90 | 35.83 |
| 34.31 | 34.26 | 36.21 |
| 33.00 | 34.70 | 36.68 |
| 32.00 | 35.16 | 37.16 |
| 31.00 | 35.61 | 37.63 |
| 30.46 | 35.86 | 37.90 |
| 30.12 | 36.00 | 38.05 |
| 29.77 | 36.19 | 38.25 |
| 29.45 | 36.33 | 38.40 |
| 29.14 | 36.50 | 38.58 |
| 28.82 | 36.65 | 38.73 |
| 28.51 | 36.82 | 38.92 |
| 28.21 | 36.98 | 39.09 |
| 27.92 | 37.14 | 39.2 |
| 27.63 | 37.31 | 39.43 |
| 27.35 | 37.46 | 39.60 |
| 27.08 | 37.62 | 39.76 |
| 26.80 | 37.78 | 39.93 |
| 26.02 | 38.26 | 40.43 |
| 25.04 | 38.91 | 41.13 |
| 24.80 | 39.08 | 41.30 |
| 22.66 | 40.73 | 43.05 |
| 20.85 | 42.40 | 44.81 |
| 19.30 | 44.05 | 46.55 |
| 17.96 | 45.72 | 48.33 |
| 16.77 | 47.41 | 50.01 |
| 14.79 | 50.86 | 53.76 |
| 12.52 | 56.16 | 59.36 |
| 10.15 | 64.16 | 67.82 |
| 6.36 | 89.66 | 94.77 |

On one works the dried slurry had the composition of 76.3 per cent. CaCO_3 and the coal used analysed:—

C=71.6; H=5.0; S=1.1; Ash=17.3; N=1.4 per cent.

From these data the table on p. 1.2 was calculated:—

Column (1) gives the fuel consumption expressed in tons of standard coal (each lb. of which, when dried, has a calorific value of 12,600 B.Th.U.'s per lb.) per 100 tons clinker made. *Column (2)* gives the corresponding theoretical volume of CO_2 as calculated. *Column (3)* gives the corrected volume of CO_2 , being the results of column (3) multiplied by the factor 1.057.

Examples of Application

Example 1.—The analysis of the exit gases on a certain works was

$$\begin{array}{r} \text{CO}_2 = 24.62 \text{ vols.} \\ \text{O}_2 = 1.95 \text{ ,,} \\ \text{N}_2 = 73.43 \text{ ,,} \\ \hline 100.00 \text{ vols.} \end{array}$$

What was the fuel consumption of the kiln?

Step 1.—Calculate the volume of air present in the exit gas.

1 volume of free oxygen corresponds to 4.762 volumes of air.

Hence 1.95 volumes of oxygen correspond to $1.95 \times 4.762 = 9.3$ volumes of air.

Step 2.—Subtract the volume of air from the volume of exit gas.

$100 - 9.3 = 90.7$ volumes of air-free gas composed of 24.62 volumes CO_2 , and $90.7 - 24.62 = 66.08$ volumes of nitrogen.

So that 66.08 volumes of nitrogen contain 24.62 volumes of CO_2 , or 100 volumes of nitrogen contain $\frac{24.62 \times 100}{66.08} = 37.2$ volumes of CO_2 .

Referring to our table, column (3), this shows a fuel consumption of 31.9 tons of coal per 100 tons of clinker made.

Found by actual large-scale measurement, 31.4 tons.

Example 2.—The exit gas of another works had the composition:—

$$\begin{array}{r} \text{CO}_2 = 23.34 \text{ vols.} \\ \text{O}_2 = 3.38 \text{ ,,} \\ \text{N}_2 = 73.28 \text{ ,,} \\ \hline 100.00 \text{ vols.} \end{array}$$

3.38 volumes of O_2 correspond to $3.38 \times 4.762 = 16.1$ volumes of air. Therefore, $100 - 16.1 = 83.90$ volumes of air-free exit gas, containing CO_2 equal to 23.34 volumes, and N_2 equal to $83.90 - 23.24 = 60.56$ volumes.

So that 60.56 volumes N_2 contain 23.34 volumes of O_2 .

So that 100 volumes N_2 contain 38.55 volumes CO_2 .

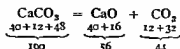
By the table, 38.55 volumes CO_2 correspond to 29.1 tons of coal per 100 tons of clinker. Found by actual large-scale trial, 29.0 tons.

Method 2.—Air consists of 21 volumes of oxygen mixed with 79 volumes of nitrogen. When coal burns, the oxygen is absorbed and nitrogen passes on. Consequently, the volume of air-free nitrogen passing out of the kiln measures the amount of oxygen absorbed and so the fuel consumed. This enables us to calculate the fuel consumption, knowing the chemical analysis of the fuel, the chemical composition of the cement clinker, and the volume analysis of the exit gas.

The method is best explained by practical examples.

By analysis a cement clinker was found to contain 65 per cent of CaO (plus MgO) So that 1 kg of cement contains 0.65 kg of CaO (plus MgO)

Now, 56 parts of CaO correspond to 44 parts of CO_2 , as is evident from the equation —



Consequently, 1 part of CaO corresponds to $44/56$ parts of CO_2 . So that 0.65 kg of CaO corresponds to $\frac{0.65 \times 44}{56} = 0.51$ kg of CO_2 , i.e., 1 kg of clinker corresponds to 0.51 kg of CO_2 .

But 1 cub metre of CO_2 at 0°C and 760 mm weighs 1.9632 kg. So that 1 cub metre of CO_2 is liberated in the exit gases to every $1.9632/0.51 = 3.85$ kg clinker produced in the furnace.

Also from the equation $\text{C} + \text{O}_2 = \text{CO}_2$ it will be seen that 44 kg of CO_2 are produced by the combustion of 12 kg of pure carbon. That is, 1.9632 kg of CO_2 are evolved from 0.536 kg of pure carbon, which is contained in 0.536/0.7115 = 0.753 kg coal.

That is, 1 cub metre of CO_2 (1.9632 kg) in the exit gases correspond to the combustion of 0.753 kg of coal of the given composition.

Now, if the coal above mentioned were completely burnt in air, the composition of the resulting exit gas can be calculated to be 18.82 volumes of CO_2 and 81.18 volumes of N_2 (total 100 volumes).

But from the equation $\text{C} + \text{O}_2 = \text{CO}_2$ it will be seen that 1 volume of CO_2 produced corresponds to 1 volume of O_2 united with carbon in the coal.

So that to every 81.18 cub metres of N_2 in the exit gases there must have entered the kiln and united therein with the carbon in the coal some 18.82 cub metres of O_2 —that is, to every 1 cub metre of N_2 issuing from the kiln we must have had $18.82/81.18 = 0.2318$ cub metres of O_2 uniting with the carbon of the coal and producing the same volume of CO_2 . With this data we are now in a position to calculate the fuel composition of the kiln by analysis of the exit gases.

Example 1—Theoretical Combustion—The exit gas from the rotary kiln had the composition —

$$\begin{array}{r} \text{CO}_2 = 26.94 \text{ vols} \\ \text{N}_2 = 73.06 \text{ „} \\ \hline 100.00 \text{ vols} \end{array}$$

Now, as just seen, to 1 cub metre of nitrogen issuing from the kiln there must have combined with the coal 0.2318 cub metre of O_2 . Therefore, to 73.06 volumes of N_2 issuing from the kiln there must have combined with the coal $73.06 \times 0.2318 = 16.94$ volumes of O_2 . And since each volume of O_2 which united with the carbon of the coal produced an equal volume of CO_2 ($\text{C} + \text{O}_2 = \text{CO}_2$), then obviously 16.94 volumes of CO_2 in the 100 volumes of exit gas must have been derived from the carbon of the coal.

But the total volume of CO_2 in 100 volumes of exit gas was 26.94. Therefore $26.94 - 16.94 = 10.00$ volumes of CO_2 must have come from the dried slurry.

Hence we have

$$\begin{array}{l} 16.94 \text{ cub metres of } \text{CO}_2 \text{ are derived from the coal} \\ 10.00 \text{ cub metres of } \text{CO}_2 \text{ are derived from the slurry} \end{array}$$

Total, 26.94 cub metres of CO_2

But 1 cub. metre CO_2 from the coal corresponds to 0.753 kg. coal; therefore, 16.94 cub. metres CO_2 from the coal correspond to $16.94 \times 0.753 = 12.76$ kg. coal. And 1 cub. metre CO_2 from slurry corresponds to 3.85 kg. of clinker; therefore, 10.00 cub. metres CO_2 correspond to $10 \times 3.85 = 38.5$ kg. of clinker.

So that 38.5 kg. clinker are produced by the combustion of 12.76 kg. of coal, or 100 tons of clinker are produced by the combustion of $\frac{12.76 \times 100}{38.5} = 33.1$ kg. coal.

Example 2.—Combustion with Excess of Air.—The exit gas from a rotary kiln had the composition:—

| | | |
|-------|-------|---------------|
| 22.81 | vols. | CO_2 |
| 3.29 | „ | O_2 |
| 73.90 | „ | N_2 |

100.00 vols.

Air has the volume composition $\text{O}_2 : \text{N}_2$ equal to 21 : 79. So that 1 volume of free O_2 is associated with $79/21 = 3.76$ volumes N_2 . The air in the exit gas is a diluent, and the composition of the exit gas free from air must be calculated. Since air has the composition above stated, it follows that 3.29 volumes of O_2 are associated with $3.29 \times 3.76 = 12.37$ volumes of N_2 .

Hence the volume of nitrogen associated with the burning processes in the kiln is $73.90 - 12.37 = 61.53$ volumes, the rest of the nitrogen belonging to the free air present.

Now, as just proved, to 1 cub. metre of air-free nitrogen issuing from the kiln there must have combined with the coal 0.2318 cub. metre of O_2 . Hence to 61.53 cub. metres of N_2 issuing from the kiln there must have combined with the coal $61.53 \times 0.2318 = 14.26$ cub. metres O_2 , which have given rise to the same volume, viz., 14.26 metres of CO_2 . So that out of the total 22.81 cub. metres of CO_2 some 14.26 cub. metres are derived from the combustion of the coal, and the residual $22.81 - 14.26 = 8.55$ cub. metres of CO_2 are derived from the slurry.

As before, the 8.55 cub. metres CO_2 from the slurry correspond to $8.55 \times 3.85 = 32.92$ kg. clinker, and 14.26 cub. metres CO_2 from the combustion of the coal correspond to $14.26 \times 0.753 = 10.74$ kg. coal burnt.

So that 32.92 kg. clinker are derived from 10.74 kg. coal burnt, or 100 kg. clinker are produced from $\frac{10.74 \times 100}{32.92} = 32.7$ kg. coal.

Example 3.—Incomplete Combustion with Production of CO.—Suppose that the exit gas has the composition:—

| | | | |
|---------------|---|------|-------|
| CO_2 | = | 26.6 | vols. |
| CO | = | 0.9 | „ |
| N_2 | = | 72.5 | „ |

100.0 vols.

There is no free oxygen present; consequently, all the nitrogen present measures the oxygen absorbed by combining with the carbon of the coal. If the combustion of the coal had been complete, there would (as before shown) have been 0.2318 cub. metre of O_2 absorbed by the coal for every cubic metre of nitrogen, or the volume of oxygen absorbed under these circumstances from the combustion of the coal is $72.5 \times 0.2318 = 16.81$ volumes of O_2 . But the combustion is not complete. There is 0.9 volume CO produced, which corresponds to 0.45 volume O_2 absorbed ($2\text{C} + \text{O}_2 = 2\text{CO}$). So that the oxygen absorbed in the complete combustion is

1 vol. 2 vols.
 $16.81 - 0.45 = 16.36$ cub. metres of O_2 .

Since the CO_2 arising from the complete combustion of the coal, plus the CO_2 evolved from the slurry, equals 26.6 volumes, it follows that the volume of CO_2 contributed by the slurry is $26.6 - 16.36 = 10.24$ cub. metres of CO_2 corresponding to $10.24 \times 3.85 = 39.42$ kg clinker. And the gases contributed by the coal are —

$16.36 \text{ CO}_2 + 0.9 \text{ CO} = 17.26$ cub. metres, corresponding to $17.26 \times 0.753 = 13.0$ kg coal (since either 1 cub. metre of CO_2 or of CO correspond to 0.753 kg of coal).

So that 39.42 kg clinker are produced by combustion of 13.0 kg coal, or 100 kg clinker are made from $\frac{13.0 \times 100}{39.42} = 33.0$ kg of coal.

Of course, it must be recollected that the figures of the coal thus calculated refer to dry coal of the composition stated.

For commercial purposes the coal can be calculated back in terms of wet, raw coal fed to the driers.

The preceding method of calculation appears to have been first suggested by Dr Albert Heiser (*Ton Industrie Zeitung*, 1913, 37, Nr 32, p 424).

Method 3 — Dr Hans Kuhl (*Zement und Mörteltechnische Studien*, I, Berlin, 1913, p 423, *Ton Industrie Zeitung*, 1913, Nr 32, p 423) has worked out a general formula for deducing the percentage fuel consumption from the complete chemical analysis of the coal, dried slurry, and exit gases.

Let x = the number of parts by weight of dry coal consumed per 100 parts by weight of clinker made

b = the percentage of carbon in the dried coal

c = the percentage of hydrogen in the dried coal

d = the percentage of nitrogen in the dried coal

e = the percentage of oxygen in the dried coal

f = the percentage of sulphur in the dried coal

g = the percentage of ash in the dried coal

m = the percentage of CaCO_3 (plus MgCO_3) in the dried slurry

n = the percentage loss by ignition of the dried slurry

$a = 0.266$ the ratio by volume of oxygen to nitrogen in the air (21 : 79)

p = the volume per cent of the CO_2 in the exit gases

q = the volume per cent of the O_2 in the exit gases

t = the volume per cent of the CO in the exit gases

$s = 100 - p - q - t$, is the volume per cent of N_2 in the exit gases

Then there are three constants, K_1 , K_2 , K_3 , which depend upon the chemical composition of the slurry from which the clinker is made, and upon the chemical composition of the coal used.

These constants may be regarded as constants for a given works using a standard type of coal and a slurry of constant composition, and they may be determined once for all and used in all subsequent calculations.

The constant K_1 depends solely upon the chemical composition of the dried slurry —

$$K_1 = \frac{400m}{100 - n} \quad (1)$$

where m is per cent CaCO_3 (plus MgCO_3) in dried slurry

n is per cent ignition loss of dried slurry

The constant K_2 depends solely upon the chemical composition of the coal used in burning —

$$K_2 = \frac{b}{3} + c + a \frac{d}{7} - \frac{e}{8} + \frac{f}{8} \quad (2)$$

where b is per cent C, c is per cent H, a is 0.266 (ratio volume O : N in air),

d is per cent. N; e is per cent. O; f is per cent. S—all referring to the dried coal.

The constant K_3 depends both upon the composition of the coal and the composition of the slurry:—

$$K_3 = \frac{b}{3} - \frac{mg}{2500 - 25n} \quad (3)$$

where b is per cent. C in dried coal; m is per cent. CaCO_3 (plus MgCO_3) in dried slurry; g is per cent. ash in coal; n is per cent. ignition loss of slurry.

There are also *two* variables, V and W , which depend upon the constantly changing composition of the exit gas:—

$$V = a \cdot s - q + \frac{t}{2} \quad (4)$$

where $a = 0.266$ (volume ratio O : N in air); s = per cent. N by volume in exit gases; q = per cent. O_2 by volume; t = per cent. CO by volume.

$$W = p + t \quad (5)$$

p is per cent. CO_2 by volume; t is per cent. CO by volume.

Now it can be proved that knowing these constants K_1 , K_2 , K_3 , and the variables V and W , then we can obtain the percentage fuel consumption, x , of the kiln as follows:—

Step 1.—Calculate the value of the constants K_1 , K_2 , K_3 from the formulæ (1), (2), and (3).

Step 2.—Calculate the value of the variables V and W from formulæ (4) and (5).

Step 3.—Substitute these values of K_1 , K_2 , K_3 , V and W in the formula:—

$$x = \frac{K_1 V}{K_2 W - K_3 V},$$

and we obtain the required value x of the number of parts by weight of dried coal (of the composition b per cent. C, c per cent. hydrogen, d per cent. nitrogen, e per cent. oxygen, f per cent. sulphur, g per cent. ash), which will produce 100 parts by weight of clinker (of composition m per cent. CaCO_3 (plus MgCO_3) in the dried slurry, n = percentage loss by ignition of the dried slurry).

Example of Application of Kuhl's Formula.—In a particular case the analysis of the raw material and coal used on the works led to the numerical values:—

$$\begin{aligned} K_1 &= 472.7. \\ K_2 &= 28.64. \\ K_3 &= 24.73. \end{aligned}$$

Also the analysis of the exit gases led to the values:—

$$\begin{aligned} \text{CO}_2 \text{ per cent.} &= p = 27.6 \text{ vols.} \\ \text{O}_2 \text{ per cent.} &= q = 0.2 \text{ ,,} \\ \text{CO per cent.} &= t = 0.2 \text{ ,,} \\ \text{N}_2 \text{ per cent.} &= S = 72.0 \text{ ,,} \\ &\hline &100.0 \text{ vols.} \end{aligned}$$

Whence

$$\begin{aligned} V &= a \cdot s - q + \frac{t}{2} \\ &= 0.266 \times 72.6 - 0.2 + 0.1 = 19.2. \\ W &= p + t \\ &= 27.6 + 0.2 = 27.8. \end{aligned}$$

Hence the weight of coal used per 10 tons of clinker is

$$\begin{aligned}
 x &= \frac{K_1}{K_2 W - K_3 V} \\
 &= \frac{472.7 \times 19.2}{28.64 \times 27.8 - 24.73 \times 19.2} \\
 &= \frac{9076}{796 - 475} \\
 &= 28.3 \text{ per cent dry coal per 100 tons clinker made}
 \end{aligned}$$

For the deduction of Kuhl's formula, reference should be made to the original paper

The difficulties of applying Kuhl's formula lie in the fact that accurate analyses of the coal and clinker are not always to hand. On the occasions on which the writer has applied the formula and checked them against direct weighed measurements of the coal and clinker, he has found that the results are too low, sometimes by several per cent. This, no doubt, is due to the causes enumerated under the first method explained above. No doubt the difficulty could be got over by calibration, and then correcting the calculated constants to fit the practical results.

Other Methods—Formulae have been worked out by Nilsson (*Ton Industrie Zeitung*, 1905, p. 1877, 1911). Also by Helbig (*Ton Industrie Zeitung*, 1906, p. 190; 1913, p. 531).

Helbig's work is interesting, as he deduces a connection between the calorific value of the fuel and the clinker output.

CHAPTER II

THE CORRECT AIR SUPPLY FOR CEMENT ROTARY KILNS

*(Tons in every instance are British tons of 2,240 lbs. To convert to
tons of 2,000 lbs. multiply by 1.12.)*

Summary.—Visual control of burning, generally practised, results in fluctuating air supply. Instrumental control is practicable, feasible, and desirable. Tables and graphs can be computed for each installation, based on chemical facts and instrumental results, which will make possible scientific control of the air supply, thus insuring maximum possible efficiency.

Fuel economy in any branch of industry is usually applauded in theory but vehemently objected to in practice, for reasons which are many and complex, but which in the end are based on human frailty.

It is a startling commentary on this that a properly designed and properly worked cement rotary kiln employing the wet process (40 per cent. slurry) can produce 100 tons of cement clinker by the combustion in the kiln of 22 tons of standard coal of 12,600 B.Th.U.'s per lb. with a good normal output.

As the kilns are invariably improperly designed (or rather not designed at all—like Topsy, they simply “grew” to their present form and dimensions) and very badly worked, very few British kilns consume less than 29 tons of standard coal, and some run up to 45 tons of standard coal per 100 tons of cement produced.

The main cause of the high fuel consumption of the rotary kiln was discovered by the writer when Director of Research of the British Portland Cement Research Association in 1925, but as the remedy meant scrapping obsolete plant and building carefully designed thermally efficient kilns, it has not been put into operation as yet, although the first steps have been taken, and some economical kilns are now running.

A minor cause of high fuel consumption was found to be the wasteful admission of excess air into the kiln. In this chapter we propose to deal with this aspect, and come on to the more complex causes in later chapters.

The interesting point of the numerous scientifically accurate kiln tests carried out between 1917 and 1925 by the British Portland Cement Research Association was the revelation for the first time of the state of hopeless inefficiency of the whole trade and the urgent necessity for reform.

Among the mass of accurate experimental data accumulated were the records of 2,139 analyses of the exit gases issuing from kilns working under all sorts of conditions.

I now propose to show how these records were utilised to discover the best amount of air to use for combustion in a rotary kiln, and how they indicate what steps should be taken in order to effect a better control than has been possible up to date.

Let us first of all study the air supply of a modern rotary kiln and show its variations from minute to minute.

Fig. 1 (L30) is a record of the variations in the exit gas analyses of the Rhoose kiln, as shown by analyses taken every few minutes over several days.

| MAXIMUM CO ₂ | | MINIMUM CO ₂ | |
|-------------------------|-----------------|-------------------------|------|
| TEST | CO ₂ | CO ₂ | CO |
| 1 | 28.4 | 0.6 | 0.34 |
| 2 | 29.0 | 0.7 | 0.1 |
| 3 | 29.2 | 1.0 | 0.0 |
| 4 | 27.4 | 1.4 | 0.0 |

| MAXIMUM O ₂ | | MINIMUM O ₂ | |
|------------------------|-----------------|------------------------|-----|
| TEST | CO ₂ | CO ₂ | CO |
| 1 | 7.8 | 19.8 | 0.0 |
| 2 | 10.2 | 6.8 | 0.0 |
| 3 | 4.6 | 20.0 | 0.0 |
| 4 | 7.6 | 18.4 | 0.0 |

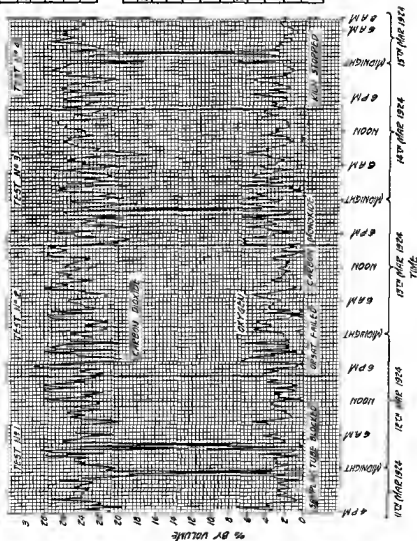
USUALLY
NIL

FIG. 1—Record of Variations in the Exit Gas Analyses taken every Few Minutes over Several Days

The composition of the gas is never for a minute steady, but varies by a series of peaks followed by valleys.

The amount of air in this gas is followed by looking at the free oxygen curve, because every 1 volume of oxygen represents about 5 volumes of air. By following this oxygen curve you will see that we hardly ever have the correct amount of air present. We either have too much or too little, and the variations from too much to too little are sudden and abrupt.

Now what is the reason of this?

You will at once see if you stand on the platform of the cement kiln and watch the burner. He uses his eye as an oxygen indicator. He first of all peers into the kiln, looking through a bit of blue glass, and he notes appearances which indicate to him that he is not supplying enough air. At once he alters the air valve, and usually now goes to the other extreme. He now lets in too much air. After a minute or two, the effect of this policy begins to show itself in minute alterations in the appearance of the white-hot material inside, and he then reverses the valve to reduce the air supply, again overshooting the mark. And so it goes on night and day in a cement kiln, a continuous variation of air supply which is never quite right.

Now I want to show you what a very bad state of affairs this is, and how we are losing valuable fuel all the time these variations in air supply are taking place. What we want is a steady flow of air and coal into the kiln, so that the composition of the exit gas does not alter appreciably from minute to minute, and under present conditions of burning I will show you that the free oxygen in the exit gas should remain constant at 1.1 per cent. by volume. With finer grinding of the coal and with more steady air control we might get below this, but things being as they are, this seems the correct percentage to use.

Admitting this for the moment, our exit gas analysis should show a steady curve when plotted against time, as shown in the dark line in Fig. 2. In studying the question of the correct air supply for a cement kiln, two main facts emerge:—

If we use too small an amount of air with our coal we obtain incomplete combustion, and a gas called carbon monoxide (CO) is produced. This means a terrible waste of fuel. To every 1 lb. of carbon which escapes in the form of carbon monoxide (CO) there is associated a direct loss of no less than 10,231 B.Th.U.'s, and an even more serious indirect loss due to the lowering of the flame temperature with consequent loss of chemical potential, represented by what we will later call the amount of high-grade available heat. For the present we will deal only with the direct heat loss, which is quite serious enough in all conscience. The main facts are shown overleaf.

Loss of Heat Due to Incomplete Combustion

Composition of one sort on coal:—

| | |
|------------|----------|
| Carbon = | 72 lbs. |
| Hydrogen = | 5 " |
| Oxygen = | 6 " |
| Ash = | 17 " |
| | <hr/> |
| | 100 lbs. |

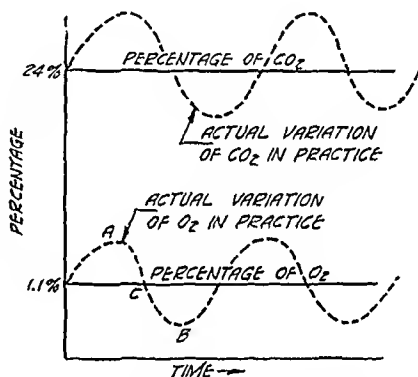
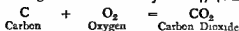


FIG. 2.—Air Supply as it should be, Steady at 1.1 per cent. Free Oxygen.

Complete Combustion (plenty of air)

1 lb of carbon burning to carbon dioxide yields 14,646 B Th U 's

**Incomplete Combustion (too little air)**

1 lb of carbon burning to carbon monoxide yields 4,415 B Th U 's



Heat Loss on Incomplete Combustion on each 1 lb of carbon is
 $14,646 - 4,415 = 10,231$ B Th U 's

On the other hand, if we use *too much air* with the coal we get complete combustion, but we now waste heat in heating an unnecessary amount of gas to a high temperature, and also we lose chemical potential by depressing the flame temperature.

So that we have, so to speak the danger of two evils—too little or too much air. Both produce a loss. What then is the right amount of air to use in the kiln?

In order to solve the problem we must go back to experiment. As I mentioned before, we have in our records some 2139 analyses of exit gases coming from cement rotary kilns. Some years ago I dissected them—and it was a terrible bit of tedious labour to do this—and the result is shown in Table 1.

In considering this table you should carefully bear in mind that the results are *statistical*. To give an example of what I mean look at column (1) and take the particular case where the free oxygen in the exit gas is shown as 0.000. Corresponding to this you will see in column (2) the carbon monoxide is returned as 0.500. Now how were these numbers obtained? In this way I went through all the exit gas analyses in which the free oxygen was returned as 0.000, and noted all the corresponding percentages of carbon monoxide. Next I added up all these percentages of carbon monoxide and divided the sum by the number of observations in which the free oxygen was put down as 0.000. I thus obtained the arithmetical average of the percentage of carbon monoxide corresponding to 0.000 per cent of oxygen in the exit gas. The number came out as 0.500 per cent of carbon monoxide when the free oxygen present in the exit gas was 0.000.

This process was repeated for all the various percentages of oxygen to be found in the exit gas analyses up to 10 per cent and in this way Table 1 was compiled.

The results are rather curious and are best exhibited graphically (Fig. 3).

You will see that the percentage of carbon monoxide rapidly rises to a maximum which is attained when the percentage of free oxygen reaches 0.1 per cent, and thereafter steadily decreases. This is quite contrary to expectation, and at first sight indicates that if we reduced the free oxygen in the gas matters would begin to improve as regards the quantity of the deleterious carbon monoxide present. This, of course is a fallacious result, and is due to the fact that we are dealing with statistical quantities.

The result is, in fact a good example of the old adage that a little knowledge is a dangerous thing. What the decrease really means is this. When the amount of air supplied becomes too small for the proper combustion of the coal (and as a consequence the amount of free oxygen in the exit gas begins to approach zero) there occur minute colour changes in the flame of the burning powdered coal, which the burner recognises. In general he is keeping a fairly careful watch on the flame, and he recognises the symptoms some time before the percentage of free oxygen actually drops to zero. He counteracts this tendency by turning on too much air. What we see, therefore, is the effect of increasing the air supply by the burner before the limit is reached. The extra air now coming into the furnace

TABLE I

Showing Connection between the Free Oxygen, Carbon Monoxide, and Loss of Fuel Due to the Carbon Monoxide in Cement Rotary Kiln Exit Gases

| (1) Average Free Oxygen in Exit Gas Per Cent. by Volume A | (2) Average Corresponding Value of CO in Exit Gas Per Cent. by Volume. B | (3) Percentage Fuel Loss Due to CO (Tons of Coal Lost per 100 Tons of Standard Coal Fired). $B \times 5.84$. |
|---|---|--|
| 0.000 | 0.500 | 2.92 |
| 0.025 | 0.699 | 4.08 |
| 0.050 | 0.994 | 5.80 |
| 0.100 | 1.703 | 9.95 |
| 0.200 | 1.195 | 6.98 |
| 0.300 | 0.862 | 5.03 |
| 0.400 | 0.630 | 3.68 |
| 0.500 | 0.480 | 2.80 |
| 0.600 | 0.323 | 1.89 |
| 0.700 | 0.214 | 1.24 |
| 0.800 | 0.198 | 1.16 |
| 0.900 | 0.134 | 0.783 |
| 1.000 | 0.109 | 0.637 |
| 1.100 | 0.081 | 0.473 |
| 1.200 | 0.0685 | 0.400 |
| 1.300 | 0.0644 | 0.376 |
| 1.400 | 0.0611 | 0.357 |
| 1.500 | 0.0563 | 0.329 |
| 1.600 | 0.0511 | 0.298 |
| 1.700 | 0.0504 | 0.294 |
| 1.800 | 0.0476 | 0.278 |
| 1.900 | 0.0469 | 0.274 |
| 2.000 | 0.0441 | 0.257 |
| 2.100 | 0.0389 | 0.227 |
| 2.200 | 0.0366 | 0.214 |
| 2.300 | 0.0318 | 0.186 |
| 2.500 | 0.0290 | 0.169 |
| 3.000 | 0.0263 | 0.154 |
| 3.500 | 0.0183 | 0.107 |
| 4.000 | 0.0149 | 0.087 |
| 5.000 | 0.0118 | 0.069 |
| 6.000 | 0.0109 | 0.064 |
| 7.000 | 0.0078 | 0.046 |
| 8.000 | 0.0046 | 0.027 |
| 9.000 | 0.0000 | 0 |
| 10.000 | 0.0000 | 0 |

starts combining with some of the carbon monoxide present, and thus diminishes its total value. In a way, therefore, the curve is a testimonial to the burners, and shows that they are vigilant.

This maximum point, therefore, is due to the human element controlling the

furnace and is not a real natural effect. In general, as the amount of oxygen decreases, the amount of carbon monoxide will increase, the curve of the carbon monoxide thus following the dotted line upwards.

Now it has been explained that in general whenever carbon monoxide is produced there is a serious loss of heat amounting to over 10,000 B Th U's per r lb. of carbon escaping as carbon monoxide (see Fig. 3).

Now, by a somewhat elaborate calculation which you will find set forth in the "Quarterly Report of the British Portland Cement Research Association for the Three Months ending 31st December 1921," it is proved that if U be the percentage volume of carbon monoxide in the exit gas, the percentage fuel loss due

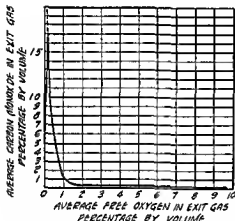


FIG. 3.—Connection between the Amount of Free Oxygen and Carbon Monoxide in Flue Gas

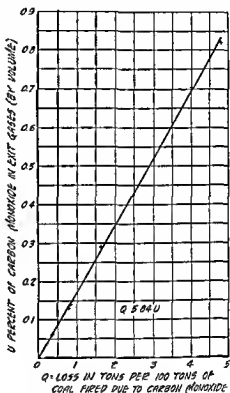


FIG. 4.—Quarterly Report of British Portland Cement Research Association showing Fuel Loss Due to Formation of Carbon Monoxide

to this carbon monoxide is 5.84 U. That is to say, out of every 100 tons of coal burnt the equivalent of 5.84 U tons are wasted owing to the formation of carbon monoxide gas during combustion of the coal. Fig. 4 shows this.

Hence we are now in a position to obtain a relationship between the amount of free oxygen in the exit gas and the loss of fuel corresponding thereto owing to the formation of carbon monoxide. Table I will make this clear.

In column (1) is the percentage of free oxygen in the exit gas.

In column (2) is the corresponding percentage of carbon monoxide.

In column (3) is the corresponding loss of fuel obtained by multiplying the percentage of carbon monoxide in column (2) by the factor 5.84, as explained above.

If we now plot column (1) against column (3), that is, the free oxygen in the

exit gas against the fuel loss due to carbon monoxide, we obtain the curve shown in Fig. 5, which gives us the percentage loss of fuel due to carbon monoxide formation corresponding to any percentage of free oxygen present in the exit gas.

In order to avoid drawing false conclusions, it should be carefully remembered that this is a *statistical result* and depends to some extent on the way the burners attend to the kiln and on the fineness to which the coal is ground.

In order that these precise numerical relationships should hold, the burners must continue to work as they worked during the tests, and the coal must continue to be pulverised in the same way.

If we alter our mode of working (*i.e.*, our statistical conditions) in general, the curves will alter to some extent, the exact amount of which can only be settled by experimenting under the new conditions.

With this general proviso we may state, then, that in general, corresponding to any given percentage of free oxygen in the exit gas there corresponds a definite average loss of fuel due to the carbon monoxide present, and that the smaller the amount of free oxygen the greater will be the loss of fuel due to carbon monoxide formation.

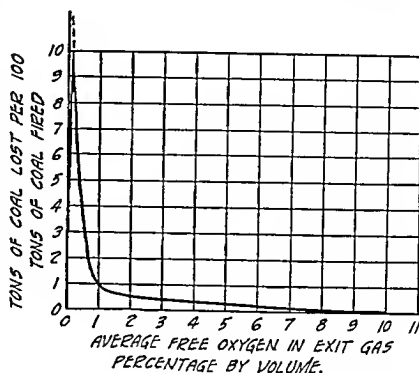


FIG. 5.—Plotting Column (1) of Table I. against Column (3) to Secure Connection between Loss of Fuel Due to Carbon Monoxide and Free Oxygen in Fuel Gas.

Loss of Fuel Due to Excess Air Present

We must now consider the loss of fuel occasioned by the air present in excess of that required for combustion of the coal. Obviously, the more air we have present in the kiln, the greater the amount of heat that will be required in heating it to the temperatures prevailing inside the kiln. Because you cannot see the air you must not consider it is not there. One cubic yard of air weighs about $2\frac{1}{4}$ lbs. and requires about the same amount of heat to raise it to a given temperature as the same weight of clinker. In other words, the specific heat of clinker is about the same as that of air. You will see, therefore, that if you use more air than is required to burn the coal, you will have more waste gas to heat up to the temperature required for clinkering, and so you will be using valuable fuel for heating this air instead of producing clinker. Moreover, the presence of this excess air lowers the flame temperature of the burning coal and thus produces a loss of what is known as chemical potential, and this leads to a serious loss of clinker, as becomes evident when we come to deal with flame temperatures. Without considering this aspect of the case, we will here merely deal with the direct loss of B.Th.U.'s which occur in the kiln, which loss may be taken as equal to the amount of heat required to heat the air from that of the external atmosphere up to the exit temperature of the gas leaving the kiln. It is quite true that the excess air when it reaches the clinkering zone or in the decarbonating zone is at a considerably higher temperature than when it leaves the kiln, but as it passes down the kiln it parts with this extra amount of heat to the walls of the kiln and the raw material it meets on the way down. Fig. 6 will make this clear.

The air enters at A at a temperature of, say, 60° F., at B it attains a temperature of about $2,800^{\circ}$ F., but going down the kiln to C, it gives up much of its heat

to the raw material, and its temperature steadily falls until it escapes at the end of the kiln at a temperature of, say, 800°F . Hence in the long run the only loss of heat is the heat required to raise the temperature of the air from 60° to 800°F as indicated at D.

You will also see from this that the higher the exit temperature, the more serious will be the loss of heat. For example, if the exit temperature was $1,000^{\circ}\text{F}$ instead of 800°F the excess air would now be heated from 60° to $1,000^{\circ}\text{F}$ instead of to 800°F , and obviously this requires more heat than before (Fig. 7).

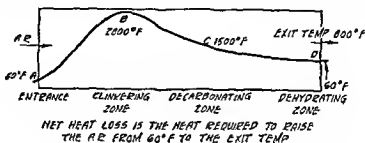


FIG. 6—How the Temperature of the Excess Air Varies in a Kiln

Notice how rapidly the loss of heat increases with the exit temperature and how important it is from an economical point of view to keep the exit temperature as low as possible. The table on p. 210 shows this in a somewhat more concrete form.

Loss of Fuel Due to the Combined Influence of the Carbon Monoxide and the Excess Air

In the preceding sections we first of all ascertained the fuel loss due to the formation of carbon monoxide considered as one factor. Then we ascertained the loss due to the excess air alone. We must now find out the combined loss due to both of these factors acting separately. You will see that we can now do this.

We have shown that corresponding to any given percentage of free oxygen in our exit gas, we know the fuel loss appertaining to the carbon monoxide and also that appertaining to the excess air present. By adding these two losses we obtain the total fuel loss due to both causes acting together. Let us give an actual example.

Let us suppose that the free oxygen in the exit gas is 1.0 per cent. Then from Table I (p. 215) we see that the loss of coal due to the carbon monoxide is 0.6370 tons of coal. Also from Table II (p. 210) we see that the fuel loss due to excess air present when the exit gas has 1.0 per cent of free oxygen present is 0.3972 tons when the exit gas temperature is 460°F .

Hence the combined fuel loss due to both causes is— $0.6370 + 0.3972 = 1.0342$ tons of coal per 100 tons of standard coal burnt in the kiln.

By repeating this calculation for the different percentage values of the free oxygen in the exit gas we arrive at the table of results on p. 211.

Fig. 8 shows the results plotted for three different exit temperatures, and you will see here again how important it is to keep the exit temperature as low as possible.

You will also notice the curious maximum point which is a statistical result due to the way the burner works the kiln by letting in excess air when the air supply becomes too small, as I explained before when dealing with the results

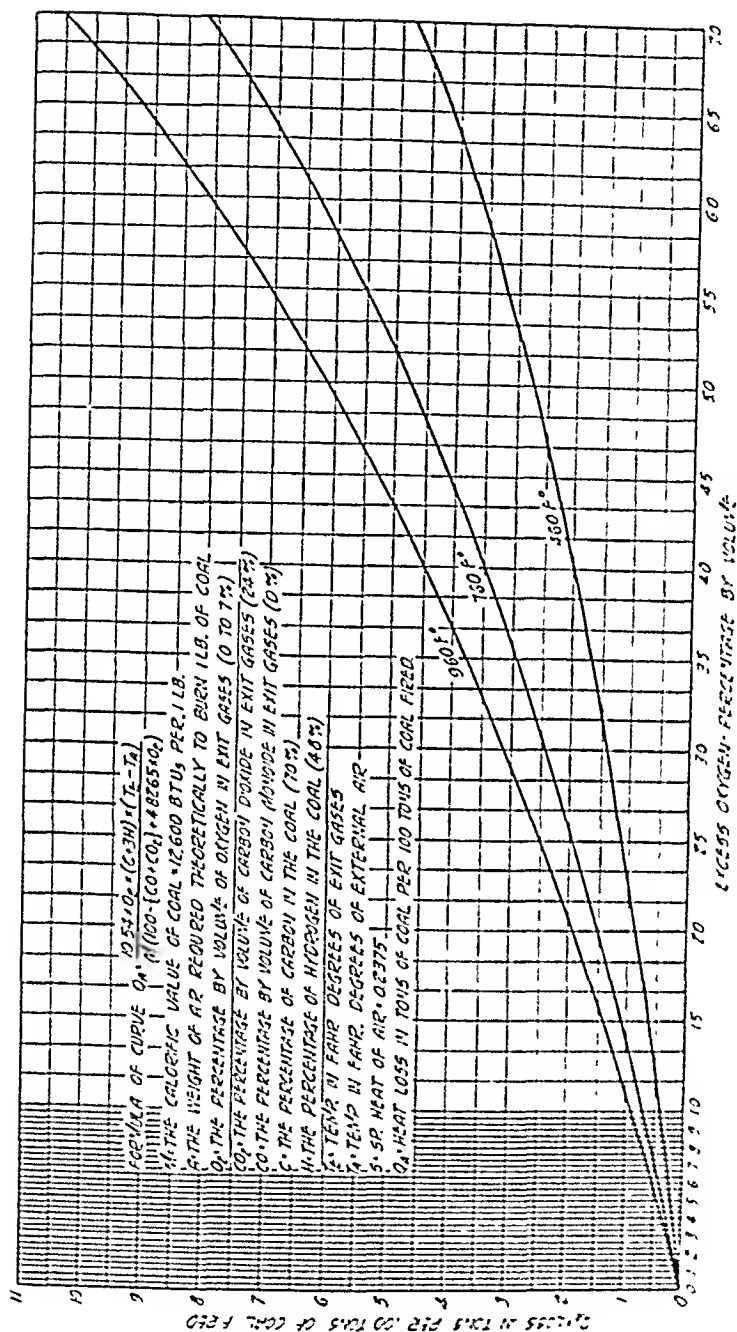


FIG. 7 - Curve showing Loss of Coal Due to Heating Excess Air. (From a Quarterly Report of British Research Bureau)

for carbon monoxide. The minimum point occurs when the percentage of free oxygen in the exit gas has the value 1.1 per cent.

We have thus arrived at the very important result that the best proportion of air to use is such as to yield an exit gas containing 1.1 per cent of free oxygen in it. It must be carefully remembered that this also is a statistical result, and depends

TABLE II

Tons of Coal Consumed in Heating Excess Air per 100 Tons of Coal Burnt in the Furnace

Standard Coal Calorific Value, 12,600 B Th U 's per Lb

| (1) Per Cent O ₂ in Volume in Exit Gases | Temperature of Exit Gases | | |
|--|---------------------------|---------------|---------------|
| | (2) 460° F | (3) 760° F | (4) 960° F |
| | Tons | Tons | Tons |
| 0.1 | 0.0364 | 0.0712 | 0.08461 |
| 0.2 | 0.0752 | 0.1321 | 0.1692 |
| 0.3 | 0.1135 | 0.1997 | 0.2560 |
| 0.4 | 0.1527 | 0.2672 | 0.3475 |
| 0.5 | 0.1919 | 0.3353 | 0.4313 |
| 0.6 | 0.2322 | 0.4055 | 0.5247 |
| 0.7 | 0.2724 | 0.4767 | 0.6129 |
| 0.8 | 0.3126 | 0.5489 | 0.7058 |
| 0.9 | 0.355 | 0.6213 | 0.7986 |
| 1.0 | 0.3972 | 0.6942 | 0.8925 |
| 1.1 | 0.4396 | 0.7267 | 0.9884 |
| 1.2 | 0.4829 | 0.845 | 1.085 |
| 1.3 | 0.5267 | 0.9214 | 1.184 |
| 1.4 | 0.5765 | 1.000 | 1.286 |
| 1.5 | 0.6221 | 1.090 | 1.400 |
| 1.6 | 0.6614 | 1.157 | 1.489 |
| 1.7 | 0.7078 | 1.239 | 1.594 |
| 1.8 | 0.7553 | 1.322 | 1.699 |
| 2.0 | 0.8512 | 1.490 | 1.916 |
| 2.5 | 1.114 | 1.949 | 2.507 |
| 3.0 | 1.377 | 2.410 | 3.100 |
| 3.5 | 1.677 | 2.927 | 3.763 |
| 4.0 | 1.993 | 3.488 | 4.483 |
| 4.5 | 2.342 | 4.097 | 5.268 |
| 5.0 | 2.774 | 4.767 | 6.130 |
| 6.0 | 3.604 | 6.304 | 8.110 |
| 7.0 | 4.685 | 8.198 | 10.55 |

upon the assumption that the kilns continue to be worked in the same way as heretofore by the burners and also that the coal continues to be ground to about the same degree of fineness. If alterations in these basic procedures take place no doubt alterations in the concomitant statistical figures will result, which would have to be redetermined by a fresh set of observations made under the new conditions.

Questions of the Steadiness of Air Supply.—Now let us go back to the question of the steadiness of the air supply and consider it in the light of the preceding remarks. You will remember that matters are so worked by the burner that he has always too much or too little air passing into the kiln, but never the exactly correct amount.

TABLE III

Total Loss of Standard Coal in Tons per 100 Tons Burnt in Furnace, Due to Combined Influence of Excess Air and Carbon Monoxide in the Exit Gases of a Cement Rotary Kiln

| Average Excess Oxygen per Cent. by Volume. | Temperature of Exit Gases. | | |
|--|----------------------------|---------|---------|
| | 460° F. | 760° F. | 960° F. |
| | Tons. | Tons. | Tons. |
| 0.0 | 2.92 | 2.92 | 2.92 |
| 0.1 | 9.98 | 10.01 | 10.02 |
| 0.2 | 7.06 | 7.11 | 7.15 |
| 0.3 | 5.15 | 5.23 | 5.28 |
| 0.4 | 3.83 | 3.95 | 4.02 |
| 0.5 | 2.57 | 2.71 | 2.81 |
| 0.6 | 2.12 | 2.30 | 2.41 |
| 0.7 | 1.53 | 1.72 | 1.86 |
| 0.8 | 1.47 | 1.70 | 1.86 |
| 0.9 | 1.14 | 1.40 | 1.58 |
| 1.0 | 1.03 | 1.33 | 1.52 |
| 1.1 | 0.91 | 1.20 | 1.46 |
| 1.2 | 0.88 | 1.24 | 1.48 |
| 1.3 | 0.90 | 1.30 | 1.56 |
| 1.4 | 0.93 | 1.36 | 1.64 |
| 1.5 | 0.95 | 1.42 | 1.73 |
| 1.6 | 0.96 | 1.45 | 1.78 |
| 1.7 | 1.00 | 1.53 | 1.88 |
| 1.8 | 1.00 | 1.60 | 1.97 |
| 2.0 | 1.10 | 1.74 | 2.17 |
| 2.5 | 1.28 | 2.12 | 2.67 |
| 3.5 | 1.78 | 3.03 | 3.87 |
| 4.0 | 2.08 | 3.57 | 4.57 |
| 5.0 | 2.80 | 4.83 | 6.19 |
| 6.0 | 3.67 | 6.36 | 8.17 |
| 7.0 | 4.73 | 8.24 | 10.57 |

Now refer to Fig. 2. When the excess oxygen is too much, as at A, he is wasting fuel, and when the excess air is too little, as at B, he is again wasting fuel. It is obvious, then, that the correct thing is for the supply of air to be steady in respect to coal so as to maintain the exit gas oxygen at about 1.1 per cent. Even if the burner arranges a series of fluctuations of air so as to maintain the average exit oxygen about right, this is not the correct thing to do, as his flame temperature will be lowered and a certain fuel wastage will occur all the same.

Hence the correct thing to do is to eliminate all fluctuations in the exit oxygen

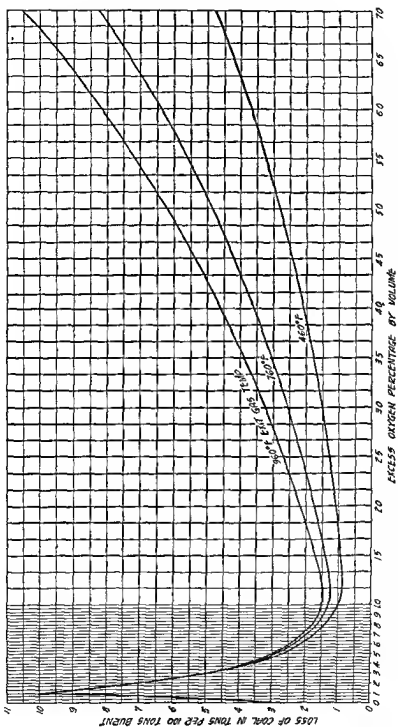


FIG. 8.—Total Loss of Standard Coal in Tons per 100 Tons Burned in Kiln Due to the Combined Influence of Excess Air and Carbon Monoxide at Three Different Exit Gas Temperatures

by ensuring a steady air supply in respect to the coal. How can we achieve this?

How to Obtain a Steady Air Supply.—In order to obtain steadiness in the air supply we must help the burner. He now depends upon his eye alone, and is surprisingly accurate. Nevertheless, you saw the fluctuations he makes. How, then, are we to help him?

Now the burner is obviously a very important man in a works. He may be likened to the driver of a motor car. What do we do with the driver of a motor car costing, say, £500? We give him all sorts of instruments to guide him. He has a little mirror to show him who is coming behind, then he is given a speedometer to tell him how fast he is going, also voltmeters, petrol gauges, brakes, horns, and all sorts of gadgets. A motor car driver, then, is well provided for. What about the driver of a large kiln costing, say, £20,000, and burning many hundreds of tons of coal a week? All the poor burner gets is a bit of blue glass,

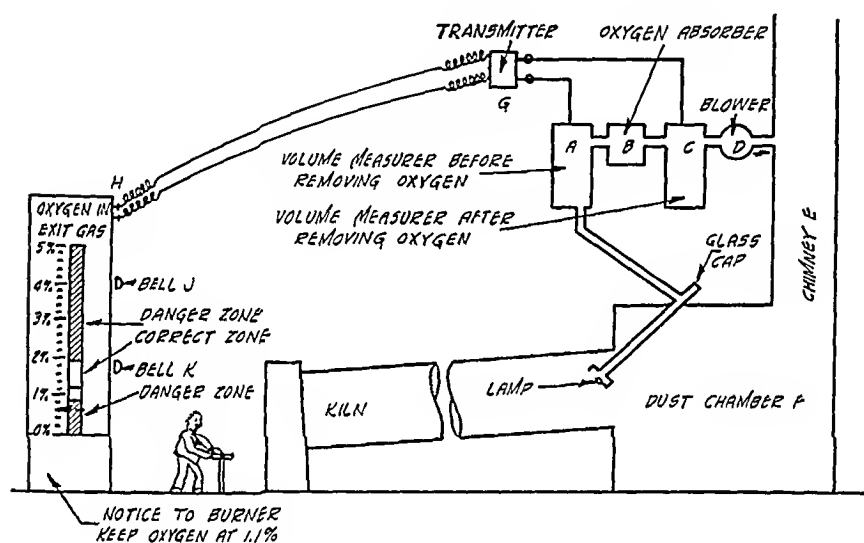


FIG. 9.—Dr Martin's Arrangement for Air Control.

which is a very crude instrument. In fact, the burner is the most neglected man on a works, whereas he should be most carefully provided with indicators for aiding him in gauging accurately the correct air supply. This brings us to the question of suitable instruments.

The present instruments in use do not reach the burner. They are kept carefully locked up in a glass case, which every now and again are inspected by a high official, but I do not think the burner regards them otherwise than with suspicion and dislike. They are no help to him, though they may indicate to the manager how in general the kiln is run.

In general what the burner wants is an instrument which indicates to him from second to second the oxygen content of the exit gases, so that without leaving his position for a moment he can at once correct his supply with a simple turn of a handle just as a driver of a motor car can regulate his speed from second to second by looking at his instruments. Fig. 9 is a diagrammatic sketch of the type of arrangement that I mean.

On the burner's platform, in a position that he can see from a distance, is an

indicator which moves up and down a scale and indicates from second to second the oxygen content of the exit gas. When the indicator gets below a position indicating less than 1 per cent of oxygen in his exit gas, a bell rings indicating danger, and the same applies when the oxygen content rises beyond 2 per cent. The job of the burner is to keep the air supply as steady as possible, so that in respect to the coal the exit gas has a uniform steady oxygen content of 1.1 per cent.

The writer has designed a recorder of this type which when fitted to an ordinary 200 ft. cement kiln will regularly save from 50 to 150 tons of coal per week—according to the degree of prevailing mismanagement—and in many cases enable an increase of output of 10 to 40 per cent per week to be attained.

APPENDIX I

Deduction of a General Formula for Calculating the Loss of Fuel Due to the Presence of Carbon Monoxide Gas in the Exit Gas for Cement Kilns—When 1 lb. of carbon is burned to carbon monoxide it produces 4,415 B Th U's, whereas when burnt to carbon dioxide it produces 14,646 B Th U's (*Kershaw*, "Fuel, Water, and Gas Analysis," 1919, p. 178).

Hence for every lb. of carbon escaping in the form of carbon monoxide there occurs the very serious loss of $14,646 - 4,415 = 10,231$ B Th U's.

Let Q = loss in tons of coal per 100 tons of coal fired owing to presence of carbon monoxide

N = the number of tons of standard coal (12,600 B Th U's per lb.) consumed in making 100 tons of clinker, carbon monoxide being present in the exit gas, *s e*, combustion is incomplete

N_1 = the standard coal consumption per 100 tons clinker which would have occurred if no carbon monoxide had been present in the exit gases, *s e*, if combustion had been perfect

P = the percentage of CaCO_3 in the dried slurry

C = the clinker output in tons per hour

X = the weight of coal fired in the kiln in tons per hour

Y = the percentage of carbon in the coal used

U = the percentage of CO by volume present in the exit gases

V = the percentage of CO_2 by volume present in the exit gases

M = the calorific value of the fuel in B Th U's per lb.

L = the loss of heat in B Th U's per lb. of carbon burnt to carbon monoxide

We take $L \approx 10,231$ B Th U's per 1 lb. carbon escaping as CO

$$\text{Then } Q = \left(1 + \frac{12P \cdot C}{(100 - 0.44P)XY} \right) \cdot \frac{L}{M} \cdot \frac{U}{U+V} \cdot \frac{Y}{1} \cdot \frac{11}{7} \quad (1)$$

When the quantities which occur in average cement rotary kilns are inserted in (1), it simplifies to the approximate formula —

$$Q \approx 5.84U \quad (1a)$$

These formulæ are deduced as follows —

Amount of Carbon Dioxide from Slurry during Clinker Formation —

From the equation $\text{CaCO}_3 \rightarrow \text{CaO} + \text{CO}_2$ it will be seen that on ignition 100 parts of CaCO_3 lose 44 parts of CO_2 , or P parts of CaCO_3 lose $0.44P$ parts of CO_2 . But 100 parts of dry slurry contain P parts of CaCO_3 , so that 100 parts of dry slurry when heated lose $0.44P$ parts of CO_2 and yield $100 - 0.44P$ parts of clinker *s e*, $(100 - 0.44P)$ tons of clinker are attended with the evolution of $0.44P$ tons of CO_2 .

Hence C tons of clinker produced are attended with the evolution from the slurry of

$$W_1 = \frac{0.44P \cdot C}{100 - 0.44P} \text{ tons of CO}_2 \quad . \quad . \quad . \quad (2)$$

W_1 represents the number of tons of CO_2 produced *per hour* in the furnace evolved from the slurry when C tons of clinker are formed.

Amount of Carbon Dioxide from Combustion of Coal.—From the equation $C + O_2 = \text{CO}_2$ it will be seen that 1 ton of free carbon completely burnt yields

$$\frac{44}{12} \text{ tons of CO}_2.$$

In 100 tons of coal there are Y tons of carbon 100 tons of coal burnt represent $\frac{44}{12}$ Y or $\frac{11}{3}$ Y tons of carbon dioxide produced.

Hence, if W_2 represent the weight in *tons* of CO_2 produced *per hour* in the kiln by the combustion of the coal, and X be the number of tons of coal fired in the kiln per hour, then

$$W_2 = \frac{11}{300} \cdot X \cdot Y \text{ tons} \quad . \quad . \quad . \quad (3)$$

Relative Weights and Volumes of Carbon Dioxide Produced from Coal and from Slurry respectively, which Occur in the Exit Gases of a Cement Rotary Kiln.—If W be the total number of tons of carbon dioxide produced in the kiln per hour, W is made up partly of carbon dioxide arising from the decomposition of the slurry during the clinker production and partly from the combustion of the coal fired.

Hence $W = W_1 + W_2$,
or $W = \frac{0.44P \cdot C}{100 - 0.44P} + \frac{11}{300} \cdot X \cdot Y \quad . \quad . \quad . \quad (4)$

Out of the total W tons of carbon dioxide in the exit gases, W_2 tons are due to coal alone.

So that for every *unit weight* of carbon dioxide in the flue there are $\frac{W_2}{W}$ units due to coal.

But the weights are directly proportional to the volumes. So that out of every 1 volume of CO_2 in the exit gasses, $\frac{W_2}{W}$ volumes are due to coal.

Since 1 volume of carbon monoxide in burning produces an equal volume of carbon dioxide, it follows that if combustion had been complete the percentage volume of CO_2 in the exit gases would have been (U + V) per cent., where U = the actual per cent. of CO, and V = per cent. of CO_2 in the exit gases, and of this the proportion due to coal alone is $(U + V) \frac{W_2}{W}$.

Hence we have the following relationship:—

$(U + V) \frac{W_2}{W}$ per cent. by volume of CO_2 in the exit gases corresponds to

$$(U + V) \frac{W_2}{W} \times \frac{12}{44} \text{ lbs. of carbon burnt,}$$

$$\text{i.e., to } (U + V) \frac{W_2}{W} \times \frac{12}{44} \times \frac{100}{Y} \text{ lbs of coal burnt,}$$

$$\text{ i.e., to } (U+V) \frac{W_2}{W} \frac{12}{44} \frac{100}{v} \text{ M B Th U's evolved}$$

where M is the calorific value of the coal in B Th U's per lb on the same proportional scale

U per cent by volume in the exit gases corresponds to U $\frac{12}{28}$ lbs of carbon burnt to CO, i.e., to a loss of L U $\frac{12}{28}$ B Th U's owing to the formation of carbon monoxide, L being the loss in B Th U's when 1 lb of carbon burns to CO instead of to CO₂

We have seen that $L = 10,731$ B Th U's

Hence out of a total possible heat evolution of

$$(U+V) \frac{W_2}{W} \frac{12}{44} \frac{100}{v} \text{ M B Th U's}$$

there are lost on account of CO formation the amount of

$$L U \frac{12}{28} \text{ B Th U's}$$

So that out of a possible 100 B Th U's evolved there are lost

$$Q = \frac{W}{W_2} \cdot \frac{L}{M} \frac{U}{U+V} \text{ i.e., } \frac{11}{7} \text{ B Th U's} \quad (5)$$

or substituting the values of W and W₂ from (3) and (4)

This reduces to

$$Q = \left(1 + \frac{12P}{(100 - 0.44P)} \right) \frac{L}{M} \frac{U}{U+V} \text{ i.e., } \frac{11}{7} \text{ B Th U's}$$

Again since 100 tons of fuel lose Q tons due to carbon monoxide formation, the loss on N tons (where N is the standard coal consumption per 100 tons of clinker produced) is $N \frac{Q}{100}$ or the corrected standard consumption is

$$N_1 = N - \frac{NQ}{100},$$

or

$$N_1 = N \frac{(100 - Q)}{100} \quad (6)$$

N₁ is the smaller standard coal consumption which would have been attained if no thermal losses due to the formation of CO had occurred i.e., if combustion had been perfect

Application of Formula to Kila Tests—The following table of data compiled from actual tests makes it possible to calculate the percentage loss of fuel Q due to the formation of carbon monoxide and also to calculate N₁, the corrected fuel consumption. The results are shown in Table III on facing page

On tabulating the percentage U of CO in the exit gases against Q, the percentage loss of fuel, we get the following table—

| | | | | | | |
|----------------------|----------|--------|--------|-------|------|-------|
| Volume per cent CO | U = 0.06 | 0.13 | 0.14 | 0.29 | 0.72 | 0.87 |
| Per cent loss fuel—Q | 0.346 | 0.7984 | 0.8394 | 1.639 | 4.01 | 4.806 |
| $\frac{Q}{U}$ | 5.80 | 6.141 | 5.996 | 5.654 | 5.57 | 5.86 |

The average value of $\frac{Q}{U}$ is 5.84 whence

$$Q = 5.84U \quad (1a)$$

TABLE III

Rotary Kiln Research

| | Symbol. | No. 1 | No. 2. | No. 3. | No. 4. | No. 5. | No. 6. |
|---|---------|--------|--------|--------|--------|--------|--------|
| Clinker output per hour, tons | C | 7.60 | 4.27 | 5.00 | 5.26 | 4.935 | 4.36 |
| Coal as fired per hour, tons | X | 2.280 | 1.221 | 1.717 | 1.590 | 1.580 | 1.246 |
| Standard coal consumption per 100 tons clinker | N | 30.7 | 29.7 | 34.1 | 32.6 | 33.8 | 30.6 |
| Calorific value of coal as fired, B.Th.U.'s per lb. | M | 12,880 | 13,108 | 12,500 | 13,590 | 13,300 | 13,500 |
| Per cent. C in coal | Y | 70.61 | 70.61 | 73.39 | 70.61 | 69.66 | 70.61 |
| CaCO ₃ in dry slurry | P | 77.6 | 76.18 | 76.2 | 76.0 | 76.1 | 76.3 |
| Analysis of exit gas per cent. by volume | V | 23.8 | 24.1 | 24.05 | 23.88 | 23.79 | 24.32 |
| CO | U | 0.13 | 0.14 | 0.82 | 0.72 | 0.29 | 0.06 |
| O ₂ | ... | 1.64 | 0.88 | 0.24 | 0.19 | 0.23 | 0.36 |

Substituting these values in the formula $Q = \left(1 + \frac{12P \cdot C}{(100 - 0.14P)XY}\right) \cdot M \cdot \frac{U}{U+V} \cdot \frac{Y}{1} \cdot \frac{11}{7}$ and $N^1 = \frac{(100-Q)}{100}$, we get the following results:—

| | Symbol. | No. 1. | No. 2. | No. 3. | No. 4. | No. 5. | No. 6. |
|--|----------------|--------|--------|--------|--------|--------|--------|
| Loss per 100 tons of coal burnt due to CO tons | Q | 0.7584 | 0.8394 | 4.806 | 4.01 | 1.629 | 0.348 |
| Standard coal consumption per 100 tons clinker if no losses had occurred | N ₁ | 30.46 | 29.46 | 32.46 | 31.3 | 33.28 | 30.5 |
| Actual standard coal consumption in presence of CO | N | 30.70 | 29.70 | 34.10 | 32.6 | 33.80 | 30.6 |
| Loss in standard coal per 100 tons clinker due to CO | tons | 0.24 | 0.24 | 1.64 | 1.3 | 0.52 | 0.1 |

This is the formula given above as applicable to cement rotary kiln practice

The loss in tons per 100 tons of coal burnt plotted against the percentage of carbon monoxide in the flue gases gives nearly, but not quite, a straight line

It will be noticed that the loss increases very rapidly with the percentage of carbon monoxide e.g., a loss of 5 per cent on firing coal occurs when 0.85 per cent CO is present in the exit gas

APPENDIX II

General Formula for Calculating the Loss of Fuel Due to Excess Air in Rotary Kilns

Let M = the calorific value of the coal used in B Th U 's per lb

A = weight of air required theoretically to burn 1 lb of coal

O_2 = per cent by volume in exit gases of oxygen

CO_2 = " " " carbon dioxide

CO = " " " carbon monoxide

N_2 = " " " nitrogen

C = " of carbon in coal

H = " of hydrogen in coal

T_E = temperature in Fahrenheit degrees in exit gases

T_A = " of external air

S = specific heat of air, here taken as 0.2375

Q_A = the number of tons of fuel per 100 tons of fuel burnt in the furnace which are wasted in heating the excess air present to the exit gas temperature

Then

$$Q_A = \frac{1054 O_2 (C + 3H) (T_E - T_A)}{M[100 - (CO + CO_2 + 4.8265 O_2)]}$$

Weight of Air Required to Burn 1 Lb of Coal — If 100 lbs of coal contain C lbs of carbon and H lbs of combustible hydrogen and if the air is taken as containing 23 per cent of oxygen by weight then from the equations



it is deduced that the weight of air required to burn C lbs of carbon is $C \frac{32}{12} \frac{100}{23}$,

and the weight of air required to burn H lbs of hydrogen is $H \cdot 8 \frac{100}{23}$

Hence the weight of air required to burn 100 lbs of coal is

$$C \frac{32}{12} \frac{100}{23} + H \cdot 8 \frac{100}{23} = 11.595(C + 3H) \text{ lbs,}$$

or the weight of air required to burn 1 lb of coal is given by

$$A = 0.11595(C + 3H) \text{ lbs} \quad (1)$$

Connection between Weight of Air Uniting with the Coal and Weight of Excess Air Supplied

From analysis —

O_2 volumes of oxygen and N_2 volumes of nitrogen are present in 100 volumes of exit gases

By Avogadro's law the weights of the volumes of two gases are in the proportion $O_2 \times 32$ to $N_2 \times 28$. Now $N_2 \times 28$ lbs. of nitrogen correspond to $(N_2 \times 28) \frac{100}{77}$ lbs. of air originally delivered, and $O_2 \times 32$ lbs. of oxygen correspond to $(O_2 \times 32) \frac{100}{23}$ lbs. of excess air originally delivered.

Hence out of $(N_2 \times 28) \frac{100}{77}$ lbs. of air originally delivered to the furnace there were $(O_2 \times 32) \frac{100}{23}$ lbs. of air in excess of that theoretically needed to burn the coal, *i.e.*, out of every 1 lb. of air originally supplied, $\frac{O_2}{N_2} \cdot \frac{32}{28} \cdot \frac{77}{23} = 3.8265 \cdot \frac{O_2}{N_2}$ lbs. are in excess of that required to burn the coal. Hence to every $(1 - 3.8265 \cdot \frac{O_2}{N_2})$ lbs. pass away with excess air, *i.e.*, to every 1 lb. of air which unites with the coal,

$$\frac{3.8265 \cdot \frac{O_2}{N_2}}{1 - 3.8265 \cdot \frac{O_2}{N_2}} \text{ lbs.}$$

of excess air pass away uselessly.

Heat Loss in Heating Excess Air.—From the preceding result it follows that since 1 lb. of coal in burning unites with $A = 0.11595(C + 3H)$ lbs. of air, the excess air present when 1 lb. of coal burns completely is

$$0.11595(C + 3H) \times \frac{3.8265 \cdot \frac{O_2}{N_2}}{1 - 3.8265 \cdot \frac{O_2}{N_2}} = \frac{0.4437 \cdot (C + 3H) \cdot \frac{O_2}{N_2}}{1 - 3.8265 \cdot \frac{O_2}{N_2}}.$$

Now, when 1 lb. of coal burns completely, there is evolved M B.Th.U.'s, and of this heat the amount used for heating the excess air is

$$\frac{0.4437 \cdot (C + 3H) \cdot \frac{O_2}{N_2}}{1 - 3.8265 \cdot \frac{O_2}{N_2}} \times S \times (T_E - T_A),$$

where S = specific heat of the air = 0.2375.
 T_E = temperature degree of exit gases.
 T_A = temperature degree of external air.

That is, out of every 100 B.Th.U.'s of heat evolved, there are wasted in heating the excess air

$$\frac{0.4437(C + 3H) \cdot \frac{O_2}{N_2} \cdot S \cdot (T_E - T_A) \cdot 100}{M \left(1 - 3.8265 \cdot \frac{O_2}{N_2} \right)},$$

Now the heat evolved is proportional to the weight of coal burnt. Hence it follows that if Q_A represents the number of tons of coal consumed in heating the

excess air out of every 100 tons of coal burnt in the furnace, we have, remembering that $N_2 = 100 - (CO + CO_2 + O_2)$, and that $S = 0.2375$ —

$$Q_A = \frac{0.4437(C+3H) O_2 (0.2375) (T_E - T_A) 100}{M \left(1 - 3.8265 \frac{O_2}{100 - [CO + CO_2 + O_2]} \right)},$$

which reduces to

$$Q_A = \frac{10.54 \times O_2 \times (C+3H) \times (T_E - T_A)}{M(100 - [CO + CO_2 + 4.8265 O_2])}$$

Conclusions — From this formula the following conclusions can be drawn —

(1) The loss of heat due to excess air is directly proportional to the excess of temperature of the exit gases over the temperature of the surrounding air. Hence the advisability of keeping the exit gases at as low a temperature as possible.

(2) The loss of heat is nearly proportional to the quantity of free oxygen present, but increases somewhat more rapidly than the percentage of free oxygen, i.e., if we double the percentage of oxygen, we somewhat more than double the loss of heat.

CHAPTER III

THE LAWS OF OUTPUT OF CEMENT ROTARY KILNS CALCULATED FROM PHYSICAL DATA

Terminal Velocities Acquired by Hot Gases Flowing through Rotary Kilns

A ROTARY kiln is a large conduit or pipe through which hot gas flows—first in an almost horizontal direction along the kiln, and then in a vertical direction up the chimney.

The force urging the gases into motion is the force of gravity. Gravity pulls downwards 1 cub. ft. of cold air outside the kiln with a greater force than it pulls downwards 1 cub. ft. of hot gas inside the kiln. In consequence of the hydrostatic pressure thus set up, the dense cold gas is pressed into the kiln so as to expel the hot gas inside up the chimney.

The case is analogous to a cork immersed in water. The cork is forced upwards owing to the fact that it is less dense than the water it displaces. That is, gravity pulls the cork downwards less strongly than it pulls downwards the surrounding water, so that the resulting hydrostatic pressure forces the cork upwards with a force equal to the difference in the weight of the cork and of an equal volume of water.

Now a cork rising upwards through the water from a great depth does not attain an infinite velocity. Its speed increases immediately after release, but the quicker it moves, the greater the resistance set up by the viscosity or internal friction of the surrounding liquid through which it moves, and very soon it attains what is known as a "terminal velocity," *i.e.*, the cork acquires a steady uniform speed upwards.

Precisely the same thing appears to happen to the hot gases rushing through the kiln. As they move they encounter resistance, partly from the walls of the kiln, but mainly owing to the viscosity or internal friction of the gases themselves.

In each rotary kiln the temperature conditions (and therefore density conditions) are more or less the same when the same process of cement manufacture is worked.

So that nearly the same limiting speed or "terminal velocity" is attained by the gases issuing from the rotary kiln.

In the case of the ordinary rotary kiln employing the wet process (40 per cent. slurry) with an exit temperature of 800° to 900° F., the limiting speed of the gases escaping from the drying zone is usually about 16.1 ft. per sec.

Percentage Fuel Consumption is Independent of Kiln Size, but Depends upon Kiln Shape

Another experimental fact which occurs in rotary kilns of the ordinary shape is this:—

A large kiln consumes sensibly the same amount of fuel per 100 tons of clinker as a small kiln of the same shape, using the same raw materials.

Kilns have been constructed ranging in size from 100 to 400 ft., and from

4 to 15 ft internal diameter, it has been found that quite small kilns (using the wet process) consume from 27 to 34 tons of coal (dry, 1 lb = 12,600 B Th U's per lb) per 100 tons of clinker, which is sensibly the fuel consumption of the large kilns

This law is not at all obvious from theoretical considerations, because in general in engineering, the larger the unit the more economical it is to run. As regards fuel consumption, this rule does not hold in rotary kilns

General Calculation of a Formula Connecting the Output of a Rotary Kiln with its Diameter

From the two preceding experimental laws it is now possible to calculate the output of clinker from a rotary kiln of any size, provided that it is of the usual shape

The calculation is carried out as follows —

Let V cub ft of exit gas at $T^\circ F$ accompany the production of 1 lb of clinker

S ft per sec be the terminal speed of the exit gas issuing from the kiln

d be the internal diameter of the kiln (brick to brick) in the calcining and drying zones, assuming that the kiln is of the same diameter in the calcining and drying zone. The bricks in this part of the kiln are sometimes 6 in thick and sometimes 5 in thick, and sometimes are coated with slurry. So that d represents the diameter inside the iron shell of the kiln less the thickness of the bricks protecting the iron shell from the hot gases

M be the output of the kiln in tons per hour

Then —

Volume of exit gas issuing from kiln is $\frac{\pi d^2 S}{4}$ cub ft per sec

Weight of clinker corresponding to this amount of slurry is

$$\frac{\pi d^2 S}{4} V \text{ lbs per sec,}$$

$$\text{or } \frac{3600}{240} \times \frac{\pi d^2 S}{4} V \text{ tons per hour}$$

$$\text{or } M = 1.63 \frac{S}{V} d^2 \text{ tons per hour} \quad (1)$$

Hence, knowing the value of the terminal speed S , the volume of gas V evolved per 1 lb clinker made, and the diameter d of the kiln we can calculate the output of the kiln under any conditions. So far as I am aware, this formula is new

APPLICATION OF FORMULA

Case I — Yields from the Wet Process Rotary Kiln

From an actual experimental determination I found that the composition of the exit gas issuing from a rotary kiln making ordinary Portland cement out of a slurry containing 40 per cent of moisture, with a fuel consumption of 33 tons of standard coal per 100 tons of cement, was

Carbon dioxide $\text{CO}_2 = 26.1$ lbs

Nitrogen, $\text{N}_2 = 47.7$ „

Steam, $\text{H}_2\text{O} = 22.9$ „

Excess air = 3.3

100.0 lbs

Also 1 lb of clinker produced under these circumstances yielded 5.36 lbs of exit gas, which, at the exit temperature of $800^\circ F$, occupied 181 cub ft. The

Here $V = 181$ cub. ft.

Substituting these values in formula (1), we have

$$M = 1.263 \times \frac{16.1}{181} d^2,$$
$$M = 0.1123 d^2 \text{ tons/hour} \quad (2)$$
$$M = 1.263 \times \frac{16.1}{133} d^2$$

$$= 0.153 d^2 \text{ tons/hour.}$$

TABLE I

| Inside Diameter of Iron Shell of Kiln Assuming Bricks 6 In. Thick in Decarbonating and Drying Zones. | (1) Inside Diameter of Kiln (Brick to Brick) in Calcining and Drying Zones. d Feet. | (2) Good Average Practice: $M = 0.1123d^2$. Clinker Output in Tons/Hour. | (3) Abnormally Good Practice: $M = 0.153d^2$. Clinker Output in Tons/Hour. |
|--|---|--|--|
| Ft. In. | Ft. In. | | |
| 2 0 | 1 0 | 0.11 | 0.15 |
| 3 0 | 2 0 | 0.45 | 0.61 |
| 4 0 | 3 0 | 1.01 | 1.37 |
| 5 0 | 4 0 | 1.80 | 2.45 |
| 6 0 | 5 0 | 2.8 | 3.8 |
| 6 4 | 5 4 | 3.2 | 4.35 |
| 6 6 | 5 6 | 3.4 | 4.6 |
| 6 10 | 5 10 | 3.8 | 5.3 |
| 7 0 | 6 0 | 4.0 | 5.5 |
| 7 6 | 6 6 | 4.8 | 6.4 |
| 7 10 | 6 10 | 5.3 | 7.1 |
| 8 0 | 7 0 | 5.5 | 7.5 |
| 8 6 | 7 6 | 6.3 | 8.6 |
| 9 0 | 8 0 | 7.2 | 9.8 |
| 9 6 | 8 6 | 8.1 | 11.1 |
| 10 0 | 9 0 | 9.1 | 12.4 |
| 11 0 | 10 0 | 11.2 | 15.3 |
| 12 0 | 11 0 | 13.6 | 18.5 |
| 13 0 | 12 0 | 16.2 | 22.0 |
| 14 0 | 13 0 | 19.0 | 25.8 |
| 15 0 | 14 0 | 22.0 | 30.6 |
| 16 0 | 15 0 | 25.3 | 34.5 |
| 17 0 | 16 0 | 28.7 | 39.2 |
| 18 0 | 17 0 | 32.4 | 44.3 |
| 19 0 | 18 0 | 36.4 | 49.7 |
| 20 0 | 19 0 | 40.5 | 55.4 |
| 21 0 | 20 0 | 45.0 | 61.3 |

Remarks on Table

Any wet process kiln possessing a smaller output than given in the third column of Table I is obviously being seriously mismanaged

The author has often found that rotary kilns are worked at less than half their proper capacity owing to lack of skilful management. The figures given in the fourth column are seldom attained with small kilns, but as the diameters increase over 10 ft the outputs given in the fourth column are often attained as the effects of slurry clinging to the brickwork and clinker rings have a less disturbing result in the case of large kilns than in the case of small kilns

Case II —Yields from Dry Process Cement Rotary Kilns

The data for applying formula (1) to dry process kilns is scanty. In one particular case the fuel consumption was 22.66 lbs of coal per 100 tons of clinker, and 1 lb of clinker evolved 3.15 lbs of exit gas at 1,900° F, which occupied approximately $V = 186.3$ cub ft. The terminal velocity of the gas was approximately $S = 20$ ft/sec.

Substituting these values for V and S in the general formula

$$M = 1.263 \frac{S}{V} d^2,$$

we get for dry process kilns

$$M = 1.263 \times \frac{20}{186.3} \times d^2 \\ = 0.1356 d^2$$

for average practice

Whence we derive the output shown in Table II on facing page

Remarks on Table

It will be noticed on comparing this table with Table I that dry process kilns yield a somewhat larger output than wet process kilns of the same diameter. Against this however must be set the cost and trouble of the preliminary drying cylinders.

Practical Importance of the Law Connecting Output with Size

The law enables us to check whether or not a kiln is being efficiently worked. By virtue of this law the author found that many kilns were only running at half their possible output, and in the case of one works he increased the output from 4 to 6 tons an hour, resulting in an increased production of over 300 tons a week, or 15,000 tons a year.

As the profits at that time were nearly £1 a ton, a saving was effected in this one works of about £15,000 per annum.

The application of the law to two very large kilns caused their output to be increased from 7 tons an hour each to 12 tons an hour each or a total weekly increase of 1,600 tons.

Per annum this means 80,000 tons extra of cement, which, at a profit of 15s per ton, meant a saving of £60,000 per annum from one works alone.

This law was applied generally to most of the kilns in Great Britain, and the resulting savings ran into some hundreds of thousands of pounds per annum.

Possibility of Increasing Output for a Given Kiln

Referring to the formula

$$M = 1.263 \frac{S}{V} d^2,$$

THE CEMENT ROTARY KILN

3.5

TABLE II

Dry Process Rotary Kilns for Cement Output in Tons per Hour from Kilns of Various Diameters

| Inside Diameter of Iron Shell of Kiln Assuming a Brick Lining 6 in. Thick in Decarbonating and Drying Zones. | Inside Diameter of Kiln (Brick to Brick) in Calcining and Drying Zones. <i>d</i> Feet. | Good Average Practice: $M=0.1356d^2$. Clinker Output in Tons/Hour. |
|--|--|---|
| Ft. | Ft. | |
| 2 | 1 | 0.14 |
| 3 | 2 | 0.54 |
| 4 | 3 | 1.22 |
| 5 | 4 | 2.17 |
| 6 | 5 | 3.39 |
| 7 | 6 | 4.88 |
| 8 | 7 | 6.64 |
| 9 | 8 | 8.68 |
| 10 | 9 | 12.20 |
| 11 | 10 | 13.56 |
| 12 | 11 | 16.40 |
| 13 | 12 | 19.52 |
| 14 | 13 | 22.9 |
| 15 | 14 | 26.6 |
| 16 | 15 | 30.5 |
| 17 | 16 | 34.7 |
| 18 | 17 | 39.2 |
| 19 | 18 | 43.9 |
| 20 | 19 | 49.0 |
| 21 | 20 | 54.2 |

where M is the clinker output in tons per hour, S is the terminal speed in feet per second of the exit gas, and V is the volume of the exit gas in cubic feet per 1 lb. clinker produced, it will be seen that the output of a given kiln may be increased in two ways:

(1) By any means which reduces the volume of gas V found per lb. of clinker made.

Consequently, reducing the fuel consumption per 1 ton clinker made, or reducing the exit temperature of the exit gas, will enable a given kiln to give a larger output.

(2) By any means which increases the terminal speed S with which the gas issues from the kiln. The introduction of fans or other means of increasing the terminal speed of the issuing gases has been patented by the writer (British Patent, 282868, 1927).

The following is a description of this patent:—

The output of clinker from a rotary kiln is proportional to the quantity of gases passing through the kiln in a given time; consequently, by increasing the linear speed with which the gases traverse the kiln, we increase the clinker output from the kiln. In the ordinary rotary kiln as operated at the present time, the gases traverse the kiln at an average linear speed not greater than 22 linear ft. per second through the clinkering and/or decarbonating zone, and at an average linear speed of about 16 linear ft. per second in the final or drying zone.

It is these conditions which limit the output of the rotary kiln to that which prevails in the usual practice at the present time

In the present patent, No 282868, this current process is deviated from inasmuch as the average linear speed of the gases traversing the kiln are maintained substantially higher than those now used in the present day practice. In this new process, on the average, the gases are maintained at an average linear speed of not less than 30 ft per second in the clinkering zone and/or decarbonating zone of the kiln, and not less than 22 ft per second in the final or drying zone, but speeds either greater or less than those mentioned may be used. By this means the output of clinker from a given kiln of the usual dimensions is increased by about 36 per cent. By causing the gases to traverse a kiln at higher average linear speeds than those mentioned, a corresponding increase of clinker output is secured, but at lower speeds a correspondingly smaller output of clinker occurs.

In one form of the invention the gases may be made to traverse all zones more rapidly than is the present practice. In another form of the invention the gases may be made to traverse some zones more rapidly than at present, but other zones at the same rate or at a slower rate than is the present practice. In a preferred form of the invention the gases may be made to traverse baffles, the gases being made to pass through the baffles at the high average linear speeds essential for an increased output. This arrangement possesses the advantage that the gases give up their heat effectively to the raw material, and at the same time internal radiation losses are prevented whereby an increased output, combined with an increased fuel economy, is secured. The average linear speed of the gases passing the baffles is measured or defined as the mean length of the sinuous path described by the greater portion of the gas passing through the baffle system divided by the time taken for a small portion of the gas to describe this path. The effect of the baffles is to make the gases passing through a kiln describe a longer path than when passing through a kiln not so fitted. So that the effect of placing baffles in a kiln is to make the kiln act as a longer kiln. This variation in the rate of travel of the gases in the different zones of the kiln may be brought about by enlarging or contracting the cross sectional area of the different zones and/or by inserting baffles therein.

One method of carrying out the invention is to place a powerful fan at the end of the kiln of such a horse power as to cause the gases to traverse the kiln at average linear speeds substantially greater than those prevailing at the present time. In order to protect the fan from the high temperatures of the issuing gases, the latter may be cooled by means of sprays and/or jets of water, and/or by passing the gases through a waste heat boiler before allowing the hot gases to impinge on the fan. Instead of a fan we may use a pump injecting cold water into the issuing gas and thus cooling it, and forcing it by injector action into the external air.

Another preferred form of carrying out the invention is to use the baffles in conjunction with a fan or other means of suction applied at the cold end of the kiln, whereby the gases are pulled through the baffles at the high average linear speeds essential for an increased output, and at the same time the baffles prevent internal radiation losses.

CHAPTER IV

THE SIZE OF THE CLINKERING ZONES OF
CEMENT ROTARY KILNS NECESSARY FOR
A GIVEN OUTPUT CALCULATED FROM
THERMODYNAMICAL DATA

Temperature Required for Making Clinker

(1) In order to make Portland cement clinker, the raw materials in the final stage must be heated to about $1,371^{\circ}\text{C}$. ($2,500^{\circ}\text{F}$.) so as to reduce the constituents to a mass not wholly fused nor wholly solid. If heated too highly, it is commonly stated that clinker becomes inert and unsuitable for making cement. If not heated highly enough, an unsound clinker is produced, which expands on cooling and falls to dust.

(2) In order to attain this clinkering temperature of the material, the brick walls of the kiln must be heated to a temperature considerably higher than this, yet not so high as to cause the brickwork to fuse.

(3) Also it is assumed that the whole of the heat necessary to make the clinker passes through the clinkering zone.

(4) Then the area of surface of the clinkering zone must be such that in a given time it must radiate away absolutely as much heat as passes through the clinkering zone per unit time. From this fact we can calculate the surface required to produce a given amount of clinker.

Calculation of the Heating Surface Required in the Clinkering Zone

Let y lbs. of clinker be made by the combustion of 1 lb. of coal of 12,600 B.Th.U.'s per lb.

M tons per hour of clinker be made = 2,240 M lbs. per hour.

In order to make this weight of clinker, the weight of coal required is $\frac{2240M}{y}$ lbs. per hour, and the B.Th.U.'s are evolved from the coal at the rate of

$$\frac{2240M}{y} \times 12,600 = \frac{M}{y} \times 28.23 \times 10^6 \text{ B.Th.U.'s per hour} \quad (1)$$

In the ordinary construction of the kiln this heat is liberated in the clinkering zone at the high temperature of over $1,371^{\circ}\text{C}$. ($2,500^{\circ}\text{F}$.). The surface exposed by the clinkering zone must be such that when heated to this temperature it will radiate away per hour as much heat as it receives. The reason of this is that the coal in burning in the clinkering zone can generate a much higher temperature than $1,371^{\circ}\text{C}$., and therefore, if there is not sufficient radiating surface exposed by the clinkering zone, the temperature of the surface of the clinkering zone will steadily increase (and thereby its rate of radiation steadily increase) until it radiates away per hour as much heat as it receives.

Now if the clinkering surface be made so large that when it attains the clinkering temperature of $1,371^{\circ}\text{C}$. ($2,500^{\circ}\text{F}$.) it radiates away as much heat in a given time as the coal imparts to it, it obviously cannot be heated beyond this temperature

by the amount of coal burnt in the kiln, and consequently overheating is impossible, because if the temperature of the brick surface rises beyond $1,371^{\circ}\text{C}$ it will radiate away more heat than it receives from the coal

Hence the area of the necessary surface of the clinking zone can be calculated on the assumption that when heated to the clinking temperature of $1,371^{\circ}\text{C}$ ($2,500^{\circ}\text{F}$) it will radiate away as much heat as the coal burnt within it can generate in the same time

Hence we have to calculate the surface S sq ft of the clinking zone, which when heated to $1,371^{\circ}\text{C}$ ($2,500^{\circ}\text{F}$), will radiate away per hour

$$28.23 \times 10^6 \times \frac{x}{y} \text{ B Th U s,}$$

where M = number of tons of clinker made per hour

y = number of lbs of clinker made by the combustion of 1 lb of coal of 12,600 B Th U s per lb

By Steffan's Law of Radiation —

$$Q = E\sigma T^4 \quad (2)$$

where Q = number of gram calories radiated per second per square centimetre of radiating surface

E = the emissivity of the brick

σ = the emissivity of a 'black body' radiation

T = the absolute temperature in centigrade degrees

According to A. T. Green* —

For brick $E = 0.72$

$$\sigma = 1.279 \times 10^{12} \text{ gram cal/sq cm}$$

Also $T = 1,371^{\circ}\text{C} = 1,371^{\circ} + 273^{\circ} \text{ absolute} = 1,644^{\circ} \text{ absolute C}$

So that $Q = 0.72 \times 1.279 \times 10^{12} \times (1,644)^4$
 $= 6.728 \text{ gram cal/s per sq cm}$

Now, converting these into the British system of units † —

$$1 \text{ sq ft} = 0.929 \times 10^3 \text{ sq cm}$$

$$1 \text{ hour} = 3.6 \times 10^3 \text{ sec}$$

$$1 \text{ B Th U} = 2.52 \text{ gram-cal}$$

Hence

$$Q = \frac{6.728}{2.52 \times 10^2} \times (3.600 \times 10^3) \times (0.929 \times 10^3)$$

or $Q = 8.93 \times 10^4 \text{ B Th U s/1 hour/1 sq ft}$

Hence the radiation from S sq ft is

$$Q = 8.93 \times 10^4 \times S \text{ B Th U s per hour} \quad (3)$$

Equating this against the heat evolved from the burning coal we have

$$\frac{M}{y} \times 28.23 \times 10^6 = 8.94 \times 10^4 \times S,$$

or $M = \frac{8.93}{28.23} \times \frac{10^6}{10^6} \times yS$

$$M = 0.3167 \times 10^{-2} yS \quad (4)$$

or $S = 315.7 \frac{M}{y} \quad (5)$

In words —

In order to make M tons of clinker per hour and assuming all the coal is burnt in the clinking zone, and that the efficiency of the kiln is such that it makes y lbs

* The Thermal Properties of Refractory Materials Trans Eng Ceramic Soc 1925 26
 vol xxv p 383

† In British Units —

$$H = \text{B Th U s/hour/square feet} = 0.7 \times 0.162 \times 10^{-3} \times 2960^4$$

of clinker for 1 lb. of standard coal burnt therein of 12,500 B.Th.U.'s per lb., then the surface necessary for the clinkering zone is given by the formula

$$S = 315.7 \frac{M}{y} \text{ sq. ft.}$$

This result is new.

Practical Application

Assume $y = 3$ lbs. of clinker made in the kiln per 1 lb. of standard coal burnt (i.e., a fuel consumption of 33 per cent., which is by no means unusual in the ordinary rotary kiln).

Then
$$S = 315.7 \frac{M}{3} = 105.2M \text{ sq. ft.} \quad (6)$$

or to every 1 ton of clinker made per hour we must allow 105.2 sq. ft. in the clinkering zone.

Hence we obtain the following table:—

TABLE I

Showing Necessary Surfaces of Clinkering Zones for Various Tonnages of Clinker from Kilns with a Fuel Consumption of 33 per Cent. per 100 Tons of Clinker

$$\text{Formula } S = 105.2M.$$

| Clinker. Tons/Hour. M. | Surface of Clinkering Zone. Square Feet. S. | Clinker. Tons/Hour. M. | Surface of Clinkering Zone. Square Feet. S. |
|------------------------------|--|------------------------------|--|
| 1 | 105.2 | 11 | 1157.0 |
| 2 | 210.4 | 12 | 1263.0 |
| 3 | 315.6 | 13 | 1368.0 |
| 4 | 420.8 | 14 | 1473.0 |
| 5 | 526.0 | 15 | 1578.0 |
| 6 | 631.2 | 16 | 1683.0 |
| 7 | 736.4 | 17 | 1788.0 |
| 8 | 841.6 | 18 | 1894.0 |
| 9 | 946.8 | 19 | 1999.0 |
| 10 | 1052.0 | 20 | 2104.0 |

Length of Clinkering Zone Appertaining to a Given Diameter and Clinker Output

In the preceding sections the correct surfaces have been calculated for a given clinker output, and a definite law expressed by the formula

$$S = 105.2M \text{ sq. ft.}$$

was found to apply.

This law, in conjunction with the law explained in Chapter III. connecting output with diameter, enables us to calculate the length of the heated surface of the clinkering zone for a kiln of any given diameter:—

Let the output be M tons of clinker per hour.

Let the diameter (reckoned brick surface to brick surface) be d ft., when the connection between M and d is given by the formula of Chapter III., viz.:—

$$M = 0.1123d^2, \text{ or } d = 2.98 \sqrt{M}. \quad (1)$$

Let L be the required length of the clinkering zone and S the necessary surface for an output of M tons an hour.

Then
$$S = \pi dL \quad (2)$$

and therefore

$$L = \frac{S}{\pi d} \quad (3)$$

Substituting

$$S = 105.2M$$

and

$$d = \sqrt{890M} = \sqrt{\frac{890}{100}} \sqrt{M},$$

or

$$d = 2.98 \sqrt{M},$$

we have

$$L = \frac{105.2M}{\pi \times 2.98 \sqrt{M}}$$

$$L = \frac{35.3}{\pi} \sqrt{M} \quad (4)$$

$$L = 11.24 \sqrt{M}$$

Hence the length L of the clinkering zone for a given clinker output M tons/hour can be easily calculated. This formula is new.

Hence we obtain the following table —

TABLE II

Showing Connection between Output and Diameter and Length of the Clinkering Zones of a Cement Rotary Kilm

| Output Tons/Hour M | Length of Clinkering Zone $L = 11.24M$ | Inside Diameter of Clinkering Zone Brick to Brick | | Inside Diameter of Iron Sheet of Clinkering Zone Allowing 9 In. Brick Lining | |
|----------------------------|--|---|----|--|----|
| | | Ft | In | Ft | In |
| 1.01 | 11.24 | 3 | 0 | 4 | 6 |
| 1.80 | 15.06 | 4 | 0 | 5 | 6 |
| 2.81 | 18.88 | 5 | 0 | 6 | 6 |
| 3.20 | 20.13 | 5 | 4 | 6 | 10 |
| 3.40 | 20.70 | 5 | 6 | 7 | 0 |
| 3.80 | 21.92 | 5 | 10 | 7 | 4 |
| 4.00 | 22.47 | 6 | 0 | 7 | 6 |
| 4.80 | 24.61 | 6 | 6 | 8 | 0 |
| 5.30 | 25.85 | 6 | 10 | 8 | 4 |
| 5.50 | 26.3 | 7 | 0 | 8 | 6 |
| 6.30 | 28.21 | 7 | 6 | 9 | 0 |
| 7.20 | 30.12 | 8 | 0 | 9 | 6 |
| 8.10 | 32.04 | 8 | 6 | 10 | 0 |
| 9.10 | 33.95 | 9 | 0 | 10 | 6 |
| 11.20 | 37.65 | 10 | 0 | 11 | 6 |
| 13.00 | 41.47 | 11 | 0 | 12 | 6 |
| 16.20 | 45.20 | 12 | 0 | 13 | 6 |
| 19.00 | 49.00 | 13 | 0 | 14 | 6 |
| 22.00 | 52.72 | 14 | 0 | 15 | 6 |
| 25.30 | 56.53 | 15 | 0 | 16 | 6 |
| 28.70 | 60.25 | 16 | 0 | 17 | 6 |
| 32.40 | 63.95 | 17 | 0 | 18 | 6 |
| 36.40 | 67.78 | 18 | 0 | 19 | 6 |
| 40.50 | 71.48 | 19 | 0 | 20 | 6 |
| 45.00 | 75.4 | 20 | 0 | 21 | 9 |

CHAPTER V

CALCULATION OF SIZE OF THE DECARBONATING ZONE FROM THERMODYNAMICAL DATA

Calculation of Number of B.Th.U.'s Radiated Away by the Walls of the Decarbonating Zone

In the Decarbonating Zone the calcium carbonate begins to decompose at 805°C . ($1,481^{\circ}\text{F}$.), and the decomposition is completed at $1,100^{\circ}\text{C}$. ($2,012^{\circ}\text{F}$.).

Hence we may take the average temperature of the material 952.5°C . ($1,747^{\circ}\text{F}$.) and the average temperature of the gas traversing this zone to be $1,126^{\circ}\text{C}$. ($2,060^{\circ}\text{F}$.).

The temperature of the walls surrounding the raw material will be lower than the temperature of the gas, but higher than the temperature of the raw material. We will, for the basis of this calculation, take the average temperature of the walls in the decarbonating zone to be $1,100^{\circ}\text{C}$. ($2,012^{\circ}\text{F}$.)

$$= 2,472^{\circ}\text{F. absolute.}$$

The absolute amount of heat which can be radiated away from the walls heated to this $T^{\circ}\text{F}$. temperature is $H = E\sigma T^4$

when E = emissivity of the brick = 0.72 .

σ = black body radiation constant

$$= 0.162 \times 10^{-8} \text{ B.Th.U.'s/hour/1 sq. ft. } 1^{\circ}\text{F.}$$

$$\text{Whence } H = 0.72 \times 0.162 \times 10^{-8} \times (2472)^4 \text{ B.Th.U.'s/1 hour/1 sq. ft.}$$

$$H = 43,358 \text{ B.Th.U.'s/1 hour/1 sq. ft.}$$

And if S be the total surface in square feet of the decarbonating zone, the total amount of heat radiated from the walls is

$$HS = 43,358S \text{ B.Th.U.'s in 1 hour} \quad . \quad . \quad . \quad (1)$$

Calculation of the Amount of Heat Absorbed by the Raw Material in the Decarbonating Zone per 1 Lb. of Clinker Made

In order to make 1 lb. of clinker we require

1.1905 lb. calcium carbonate, CaCO_3 .

0.1846 lb. pure kaolin, $\text{K}_2\text{O} \cdot \text{Al}_2\text{O}_3 \cdot 6\text{SiO}_2$.

0.1540 lb. silica, SiO_2 .

0.0312 lb. ferric oxide, Fe_2O_3 .

Total, 1.5603 lb.

In order to decompose 1.1905 lbs. of calcium carbonate at 805°C . requires $1.1905 \times 682 = 811.92 \text{ B.Th.U.'s}$.

There remains 1 lb. weight of raw materials which must be heated from $1,481^{\circ}\text{F}$. (805°C .) to $2,012^{\circ}\text{F}$. ($1,100^{\circ}\text{C}$.). The specific heat of the raw material may be taken as 0.28 , so that the quantity of heat absorbed in this operation is

$$0.28 \times (2012 - 1481) = 148.7 \text{ B.Th.U.'s.}$$

Hence the total amount of heat absorbed in the decarbonating zone per 1 lb of clinker produced is

$$148.7 + 811.9 = 960.6 \text{ B Th U 's,}$$

or per 1 ton of clinker made, the number of B Th U 's absorbed or rendered latent in the decarbonating zone is

$$960.6 \times 2240 = 2.152 \times 10^6$$

Calculation of the Surface Required in the Decarbonating Zone per M Tons of Clinker Made in the Kiln

Let y lbs of clinker be made by the combustion in the kiln of 1 lb of standard coal of 12,600 B Th U 's per lb

Let M tons per hour of clinker be made = $2,240M$ lbs per hour

In order to make this weight of clinker, the weight of coal required is $2240 \frac{M}{y}$ lbs per hour, and the number of B Th U 's evolved from the coal per hour are

$$\frac{2240M}{y} \times 12600 = 28.23 \times 10^6 \frac{M}{y} \text{ B Th U 's per hour}$$

All this heat pours down through the decarbonating zone. But in making M tons of clinker there are absorbed or rendered latent in the decarbonating zone

$$2.152 \times 10^6 M \text{ B Th U 's} \quad (2)$$

Hence there occurs an unbalanced residue of heat pouring through the decarbonating zone of

$$\begin{aligned} & 28.23 \times 10^6 \frac{M}{y} - 2.152 \times 10^6 M \text{ B Th U 's} \\ & = M \times 10^6 \left(\frac{28.23}{y} - 2.152 \right) \text{ B Th U 's} \end{aligned}$$

This heat, being free, must be counterbalanced by the same amount of heat radiated from the walls of the decarbonating zone because the temperature of the walls of the decarbonating zone must steadily increase as the free heat plays upon them until they radiate back as much heat as they receive in a given time.

Hence, equating the heat radiated away by the walls against the free heat from the coal, we get

$$43358S = M \times 10^6 \left(\frac{28.23}{y} - 2.152 \right)$$

$$\text{or} \quad S = \frac{M \times 10^6}{43358 \times 10^4} \left(\frac{28.23}{y} - 2.152 \right)$$

$$= M \times 100 \left(\frac{6.51}{y} - 0.4963 \right)$$

$$\text{or} \quad S = M \left(\frac{651}{y} - 49.63 \right) \quad (3)$$

where S = number of square feet required in decarbonating zone to make M tons of clinker per hour

M = number of tons of clinker made per hour by kiln

y = number of lbs of clinker made per 1 lb of standard coal burnt in kiln

This formula is new

Practical Application of Formula

Assume $y = 3$ lbs of clinker per 1 lb of coal burnt (i.e., a fuel consumption of 33 per cent, which is an ample allowance)

Then the surface is

$$S = M \left(\frac{651}{3} - 49.63 \right)$$

$$M = (217 - 49.63)$$

$$S = 167.4 \text{ M} \quad (4)$$

From this formula we can calculate the required surface in square feet in the decarbonating zone for any required clinker output.

TABLE I

*Showing Surface Necessary in Decarbonating Zone for
Various Tonnages of Clinker per Hour*

$$S = 167.4 \text{ M.}$$

| Tons of Clinker per Hour. M Tons. | Requisite Surface. S Square Feet. |
|---|---|
| 1 | 167.4 |
| 2 | 335.0 |
| 3 | 502.0 |
| 4 | 670.0 |
| 5 | 837.0 |
| 6 | 1005.0 |
| 7 | 1172.0 |
| 8 | 1339.0 |
| 9 | 1507.0 |
| 10 | 1674.0 |
| 11 | 1841.0 |
| 12 | 2009.0 |
| 13 | 2176.0 |
| 14 | 2344.0 |
| 15 | 2512.0 |
| 16 | 2678.0 |
| 17 | 2846.0 |
| 18 | 3013.0 |
| 19 | 3181.0 |
| 20 | 3348.0 |

Calculation of Length of Decarbonating Zones Corresponding to a Given Diameter and Clinker Output

In the preceding section the correct surface has been calculated for a given clinker output, and a definite law, expressed by the formula

$$S = 167.4M,$$

was found to apply.

This law, in conjunction with the law explained in Chapter II. connecting output with diameter, enables us to calculate the length of the heated surface of the decarbonating zone for a kiln of any given diameter.

Let the output be M tons of clinker per hour.

Let the diameter (reckoned brick surface to brick surface) be d ft.

Then the connection between M and d is given by

$$M = 0.1123d^2, \text{ or } d = 2.98 \sqrt{M} \quad (1)$$

Let L feet be the required length of the clinkering zone, and S the necessary surface for an output of M tons an hour

$$\text{Then} \quad S = \pi dL \quad (2)$$

$$\text{or} \quad L = \frac{S}{\pi d} \quad (3)$$

$$\text{Substituting} \quad S = 167.4M$$

$$d = 2.98 \sqrt{M},$$

$$\text{we have} \quad L = \frac{167.4}{\pi \times 2.98} \frac{M}{\sqrt{M}}$$

$$= \frac{167.4}{9.363} \sqrt{M},$$

$$\text{or} \quad L = 17.88 \sqrt{M}$$

Hence the length L of the decarbonating zone for a given clinker output M can be easily calculated by this new formula

The following table gives the results —

TABLE II

Giving the Connection between Output, Diameter, and Length of the Decarbonating Zones of a Cement Rotary Kiln

| Output Tons/Hour M | Length of Decarbonating Zone $L = 17.88M$ | Inside Diameter of Decarbonating Zone Brick to Brick | | Inside Diameter of Iron Shell of Decarbonating Zone Allowing for 6 In. Brick Lining | |
|----------------------------|---|--|----|---|----|
| | | Ft | In | Ft | In |
| 1.0 | 17.88 | 3 | 0 | 4 | 0 |
| 1.80 | 23.95 | 4 | 0 | 5 | 0 |
| 2.81 | 30.03 | 5 | 0 | 6 | 0 |
| 3.20 | 32.0 | 5 | 4 | 6 | 4 |
| 3.40 | 32.9 | 5 | 6 | 6 | 6 |
| 3.80 | 34.9 | 5 | 10 | 6 | 10 |
| 4.00 | 35.7 | 6 | 0 | 7 | 0 |
| 4.80 | 39.1 | 6 | 6 | 7 | 6 |
| 5.30 | 41.1 | 6 | 10 | 7 | 10 |
| 5.50 | 41.8 | 7 | 0 | 8 | 0 |
| 6.30 | 44.9 | 7 | 6 | 8 | 6 |
| 7.20 | 47.9 | 8 | 0 | 9 | 0 |
| 8.10 | 50.9 | 8 | 6 | 9 | 6 |
| 9.10 | 54.0 | 9 | 0 | 10 | 0 |
| 11.20 | 59.9 | 10 | 0 | 11 | 0 |
| 13.60 | 66.0 | 11 | 0 | 12 | 0 |
| 16.20 | 71.9 | 12 | 0 | 13 | 0 |
| 19.00 | 77.9 | 13 | 0 | 14 | 0 |
| 22.00 | 83.8 | 14 | 0 | 15 | 0 |
| 25.30 | 89.9 | 15 | 0 | 16 | 0 |
| 28.70 | 95.8 | 16 | 0 | 17 | 0 |
| 32.40 | 102.0 | 17 | 0 | 18 | 0 |
| 36.40 | 108.0 | 18 | 0 | 19 | 0 |
| 40.50 | 114.0 | 19 | 0 | 20 | 0 |
| 45.00 | 120.0 | 20 | 0 | 21 | 0 |

CHAPTER VI

THE DIMENSIONS OF ROTARY KILNS, INCLUDING RULES FOR FINDING THEIR LENGTHS AND DIAMETERS, AND SIZES OF DECARBONATING, PREHEATING, AND DRYING ZONES OF BOTH WET AND DRY PROCESS KILNS

THE size of the clinkering and decarbonating zones can be calculated from the radiatory constants alone.

In the case of the preheating zone, and drying zone, however, much of the heat-transfer takes place through convection and conduction, radiation playing a progressively smaller part as the temperature decreases.

So that theoretical deductions of their respective sizes are difficult.

However, as a matter of empirical experimenting it has been found that the following formulæ give approximately correct results in the case of the wet process employing 40 per cent. slurry moisture :—

Let L be the total length of the kiln.

Then :—

$$\text{Cooler End} = 0.0952L = \frac{2}{21}L \text{ approximate.}$$

$$\text{Clinkering Zone} = 0.1905L = \frac{4}{21}L \text{ approximate.}$$

$$\text{CO}_2\text{-Expulsion Zone} = 0.2381L = \frac{5}{21}L \text{ (say, } \frac{1}{3}L\text{).}$$

$$\text{Preheating Zone} = 0.1429L = \frac{3}{21}L.$$

$$\text{Drying Zone} = 0.3333L = \frac{7}{21}L \text{ (say, } \frac{1}{3}L\text{).}$$

| Cooler End. | Clinkering Zone. | CO ₂ -Expulsion Zone. | Preheating Zone. | Drying Zone. |
|-----------------|---------------------|-------------------------------------|---------------------|-----------------|
| $\frac{2}{21}L$ | $\frac{4}{21}L$ | $\frac{5}{21}L$ | $\frac{3}{21}L$ | $\frac{7}{21}L$ |

Ratio of Length of Kiln to Diameter

Wet Process.—If d' be the diameter in feet of the kiln, brick face to brick face, and if L feet be its length, a good rule is

$$\left. \begin{array}{l} L = 27d' \text{ minimum} \\ L = 29d' \text{ maximum} \end{array} \right\} \dots \dots \dots (1)$$

The trouble is how to determine d' . A formula which the writer has worked out and which allows for the fact that the larger the diameter the smaller the relative cooling surface prescribed, and which gives very reliable results is this :—

$$L = 25(d - 1.5) + 0.2(d - 1.5)^2 \dots \dots \dots (2)$$

when L is the length of the kiln in feet.

d is the diameter of the iron shell in the decarbonating and drying zones.

Where the diameters vary down the kiln, d must be taken as the *average* diameter.

The following table is calculated from formula (2) :—

TABLE I

Showing Connection between Diameters and Lengths of Wet Process Cement Rotary Kilns

$$L = 25(d - 1.5) + 0.2(d - 1.5)^2$$

| Diameter of Shell <i>d</i> Feet | | Length <i>L</i> Feet | Diameter of Shell <i>d</i> Feet | | Length <i>L</i> Feet |
|------------------------------------|----|-------------------------|------------------------------------|----|-------------------------|
| Ft | In | Ft | Ft | In | Ft |
| 3 | 0 | 38 | 10 | | 227 |
| 4 | 0 | 64 | 11 | | 255 |
| 5 | 0 | 90 | 12 | | 284 |
| 6 | 0 | 117 | 13 | | 314 |
| 6 | 6 | 130 | 14 | | 344 |
| 7 | 0 | 144 | 15 | | 374 |
| 7 | 6 | 157 | 16 | | 405 |
| 8 | 0 | 171 | 17 | | 435 |
| 8 | 6 | 185 | 18 | | 467 |
| 9 | 0 | 199 | 19 | | 499 |
| 9 | 6 | 213 | 20 | | 531 |

For Dry Process Cement Kilns the material is already dry before it reaches the kiln, and consequently the kiln is shorter by approximately the drying zone

The following formula gives fairly reliable results for the connection between the length in feet of a dry process kiln and its diameter —

$$L = 20(d - 1.5) + 0.2(d - 1.5)^2,$$

where *L* and *d* have the same meaning as before

The following table is compiled from this formula —

TABLE II

Showing Connection between Lengths and Diameters of Dry Process Cement Rotary Kilns

$$L = 20(d - 1.5) + 0.2(d - 1.5)^2$$

| Diameter of Iron Shell <i>d</i> Feet | | Length of kiln <i>L</i> Feet | Diameter of Iron Shell <i>d</i> Feet | | Length of kiln <i>L</i> Feet |
|--|----|---------------------------------|--|----|---------------------------------|
| Ft | In | Ft | In | Ft | Ft |
| 3 | 0 | 30 | 6 | 10 | 185 |
| 4 | 0 | 51 | 0 | 11 | 208 |
| 5 | 0 | 72 | 0 | 12 | 232 |
| 6 | 0 | 94 | 0 | 13 | 257 |
| 6 | 6 | 106 | 0 | 14 | 281 |
| 7 | 0 | 117 | 0 | 15 | 307 |
| 7 | 6 | 129 | 0 | 16 | 332 |
| 8 | 0 | 139 | 0 | 17 | 358 |
| 8 | 6 | 150 | 0 | 18 | 385 |
| 9 | 0 | 161 | 0 | 19 | 412 |
| 9 | 6 | 173 | 0 | 20 | 439 |

Other formula —

1 A rough practical rule for kiln capacity (wet process, 40 per cent slurry) is 0.02 cwt of clinker per kiln hour per cubic foot of internal capacity

2 Another rule is 0.336 lb clinker per kiln hour per 100 sq ft heating surface (wet process, 40 per cent slurry moisture)

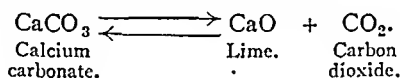
3 Harrison's formula is Number of barrels of cement clinker (each 380 lbs weight) per hour = $0.6 \times \text{inside diameter of kiln in feet} \times \text{length of kiln in feet}$ (wet process cement kilns, 40 per cent slurry moisture)

CHAPTER VII

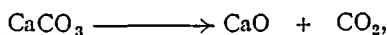
TEMPERATURES AT WHICH CALCIUM AND MAGNESIUM CARBONATES DECOMPOSE IN THE CEMENT ROTARY KILN

§ 1. THE theory of the expulsion of carbon dioxide from carbonates is well known and is explained in works on physical chemistry.

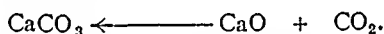
The reaction is reversible, and takes place according to the equation



Corresponding to each particular temperature T there is a definite fixed pressure p of the carbon dioxide, such that if the pressure of the carbon dioxide in the surrounding atmosphere be *less* than this pressure, the calcium carbonate will completely decompose, thus :—



whereas, if the surrounding pressure of the carbon dioxide be greater than this, the lime will be completely retransformed into CaCO_3 , thus :—



The following is a table of these dissociation pressures taken from John Johnstone's results,* which are the most accurate determinations to date, and have recently been confirmed † :—

| Temperature. | Equilibrium Pressure of Carbon Dioxide. | Temperature. | Equilibrium Pressure of Carbon Dioxide. |
|--------------|---|--------------|---|
| ° C. | Mm. | ° C. | Mm. |
| 500 | 0.11 | 800 | 168 |
| 550 | 0.57 | 850 | 373 |
| 600 | 2.35 | 898 | 760 |
| 650 | 8.2 | 900 | 773 |
| 700 | 25.3 | 950 | 1,490 |
| 750 | 68.0 | 1,000 | 2,710 |

* John Johnstone, "The Thermal Dissociation of Calcium Carbonate," *Jour. Amer. Chem. Soc.*, 1910, xxxii., 938.

† Pierre Jolibois et Bouvier, *Comptes Rendus*, 1921, 172, 1182-1183.

These pressures are connected by the formula

$$\log p = -\frac{2340}{T} + 1.1 \log T - 0.0012T + 8.882,$$

where p = the equilibrium pressure of the CO_2 expressed in millimetres of mercury,
 T = the absolute temperature expressed in $^{\circ}\text{C}$,

and are plotted in a curve, as illustrated in Fig. 1. At 898°C the equilibrium pressure of the evolved carbon dioxide is 1 atmosphere. This merely means that if the CO_2 in the surrounding atmosphere exerts a pressure slightly less than 1 atmosphere, the calcium carbonate will completely decompose if kept at 898°C for a sufficient length of time, whereas if the CO_2 pressure be slightly

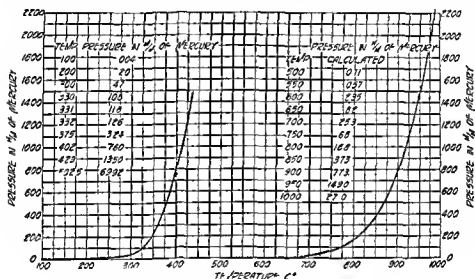


FIG. 1.—The Dissociation Pressures of Magnesium Carbonate (left) and Calcium Carbonate (right)

greater than 1 atmosphere, no decomposition ensues no matter how long the CaCO_3 be heated to 898°C .

It is obvious, therefore that the rate at which the calcium carbonate decomposes depends almost entirely upon the conditions under which it is heated, e.g., if the calcium carbonate is heated in a rapid current of gas free from carbon dioxide, so that the carbon dioxide is removed from the material almost as rapidly as evolved, the rate of decomposition will be far more rapid than if the CaCO_3 be heated in an atmosphere of gas containing much CO_2 .

Consequently, although the dissociation of CaCO_3 under atmospheric pressure in an atmosphere of CO_2 will only proceed to completion at a temperature about 898°C , nevertheless, in a rotary kiln, where the mass is heated in a current of gas of which only a percentage consists of carbon dioxide, the decomposition will proceed almost to completion at a fairly rapid rate before this temperature is attained. We can, however, calculate the temperature at which the decomposition of the CaCO_3 begins in the rotary kiln as follows—

Calculation of the Temperature at which CaCO_3 begins to Evolve CO_2 in a Cement Rotary Kiln

Assume as approximately correct in ordinary practice that the weight of CO_2 expelled from the slurry is half the weight of the clinker produced, also that 100 tons clinker are produced by burning 30 tons of coal containing 70 per cent. carbon. Then CO_2 from the slurry is 50 tons and CO_2 from the coal is 77 tons, so that 1 part of CO_2 consists of $\frac{50}{127}$ parts from slurry and $\frac{77}{127}$ from coal or 24 parts of CO_2 consist of 9.5 from slurry and 14.5 from coal.

Hence we deduce that if the *exit gas* from the kiln consists of 24 volumes CO_2 , 1 volume O, and 75 volumes N per 100 volumes, this gas, as it comes from the sintering zone and before enriched with the carbon dioxide from the chalk, will have the $\text{CO}_2=16$ per cent., $\text{O}_2=1.1$ per cent., $\text{N}=82.9$ per cent. composition by volume. Hence the *partial pressure* of the CO_2 in the furnace gases as they

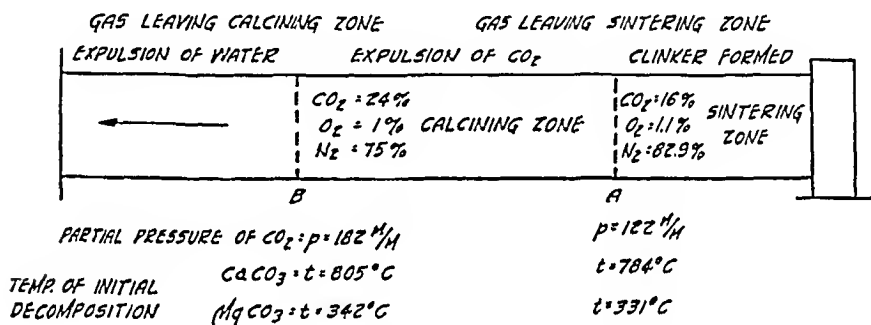


FIG. 2.—Composition of Gaseous Atmosphere Inside the Cement Rotary Kiln.

leave the sintering zone must be 0.16 atmosphere = 122 mm. of mercury, and as they leave the *calcining zone* must be 0.24 atmosphere = 182 mm. of mercury. This is illustrated in Fig. 2.

Now, referring to the curve of equilibrium pressure for the CO_2 , we see that when $p = 122$ mm., the temperature at which the chalk begins to decompose is 784°C ., whereas when $p = 182$ mm., the temperature of initial decomposition of the chalk is 805°C .

Conclusion.—In a rotary cement kiln worked under normal conditions the chalk feed begins to evolve carbon dioxide at 805°C ., but as it approached the sintering zone the temperature at which it begins to evolve CO_2 is reduced to 784°C .

Practically this means that the evolution of carbon dioxide from chalk is almost infinitely *slow* when its temperature is 805° and 784°C . respectively at the points B and A respectively in the kiln, but as the temperature of the raw material increases beyond these limits, the rapidity of expulsion is *rapidly increased*.

By the time the temperature of the chalk has reached 900°C . under these conditions the *bulk* of the carbon dioxide probably has been expelled, the last traces being eliminated probably at $1,100^\circ \text{C}$.

It should be observed that the *time factor* is of great importance. For example,

years would be required for the complete expulsion of the carbon dioxide from the chalk heated to 805°C at the point *n* of the furnace in Fig 2, where it is exposed to an atmosphere consisting of $\text{CO}_2 = 24$ per cent, $\text{O} = 1.0$ per cent, $\text{N} = 75$ per cent. Whereas, if the chalk be heated in the same atmosphere to, say, 900°C and maintained thereat, probably a few minutes will suffice. Unfortunately, the exact times required for decomposition are quite unknown. Indeed, it would be very desirable to determine the *minimum time* of passage of chalk through the calcining zone of a furnace maintained at different temperatures, say, 800° to $1,100^{\circ}\text{C}$, in order to arrive at this constant for the different temperatures. The influence of the varying speeds of the kiln gases over the surface of the hot chalk would also have to be investigated at the same time.*

Having collected such data, one would be then in a position to design a kiln on scientific lines so as to work it for a maximum possible theoretical output.

At the present time kilns are designed on more or less rule of thumb methods founded on practical experience, as the engineers who design such kilns have never been supplied with the chemical and physical data whereon to base their dimensions and speed of passage.

Calculation of the Temperature of Decomposition of Magnesium Carbonate in the Cement Rotary Kiln

In the case of magnesium carbonate, Marc and Simek† determined the dissociation pressures of magnesium carbonate as follows —

| Temperature | Dissociation Pressure of CO_2 | Temperature | Dissociation Pressure of CO_2 |
|--------------------|--|--------------------|--|
| $^{\circ}\text{C}$ | Mm | $^{\circ}\text{C}$ | Mm |
| 100 | 0.04 | 332.0 | 126 |
| 200 | 2.00 | 375.0 | 524 |
| 300 | 47.00 | 402.0 | 760 = 1 atmos |
| 330 | 108.00 | 429.0 | 1 315 = 1.73 , |
| 331 | 118.00 | 502.6 | 6 992 9.2 , |

The pressures are connected by the formula

$$\log p = \log B + N \log T,$$

where p = pressures in atmospheres

$$B = -47.271$$

$$N = 16.70$$

The curve is plotted in Fig 1

Taking as before the partial pressure of the carbon dioxide in the furnace gases as ranging from 122 mm at point *A* (end of sintering zone) to 182 mm at *B* (end of calcining zone), we get the corresponding temperatures of initial decomposition of the magnesium carbonate as 331°C at point *A* and 342°C at point *B*.

* See *C. C. Furnas* The Rate of Calcination of Limestone Rock Products 20th June 1931 Vol xxxiv p 32

† *Robert Marc and A. Simek* The Thermal Dissociation of Magnesium Carbonate *Zeitsch. Anorg. Chem.*, 1913 82 1749

At these two temperatures, then, in the kiln and at the two places the evolution of CO_2 from the MgCO_3 is infinitely slow. When we heat the MgCO_3 above these temperatures the evolution of gas occurs at a rapidly increasing rate.

It has been found, however, that the *rate* at which the carbon dioxide is expelled from the magnesium carbonate is *very slow* below 600°C .

For example, *Grunberg** found that complete decomposition of the magnesium carbonate in a finite time occurred when heated in a stream of air and in a stream of carbon dioxide at $470 \pm 15^\circ \text{C}$. and $650 \pm 20^\circ \text{C}$. respectively.

According to *C. Schubert*,† MgCO_3 begins to decompose at 350°C ., but the process is only complete at 900°C .

APPENDIX

For completeness' sake there are collected together the following references on this subject which have been published in the literature relating to cement:—

W. B. Newbury (*Cement and Engineering News*, 1912; fully reported in the *Ton.-Industrie Zeitung*, 1902, 26, ii., 1215) states that the loss of combined water and organic matter is marked at 840°C . ($1,544^\circ \text{F}$.). At 870°C . ($1,598^\circ \text{F}$.) much carbon dioxide has been expelled.

Dr Bruhn (*Ton.-Industrie Zeitung*, 1903, 27, i., 235) states that the loss of carbon dioxide from the carbonate only proceeds very slowly at 530° to 600°C . ($1,069^\circ$ to $1,112^\circ \text{F}$.), and requires a temperature at 800° to $1,000^\circ \text{C}$. ($1,472^\circ$ to $1,832^\circ \text{F}$.) in order to proceed rapidly to completion.

E. C. Seyer ("Report on a Test on a Portland Cement Plant," *Engineering News*, 1905, liv., No. 25, p. 664) gives the temperature at which the evolution of carbon dioxide becomes active as 540°C . ($1,000^\circ \text{F}$.). This must be considerably too low.

H. S. Spackman (*Ton.-Industrie Zeitung*, 1906, 30, i., 847) states that the temperature at which the main amount of carbon dioxide is expelled is 812°C . ($1,494^\circ \text{F}$.), this process being completed by the time the temperature of the mass has attained $1,000^\circ \text{C}$. ($1,832^\circ \text{F}$.).

Dr Brendall (*Ton.-Industrie Zeitung*, 1908, 32, 1084) puts the temperature at which carbon dioxide begins to be rapidly evolved in the rotary kiln as 750°C . ($1,382^\circ \text{F}$.), the process being practically finished at 950°C . ($1,742^\circ \text{F}$.).

J. H. Schutt (*Ton.-Industrie Zeitung*, 1913, 37, 2037; 1914, 38, 1021) assumed that the temperature at which carbon dioxide is expelled in the rotary kiln lies between 750°C . ($1,382^\circ \text{F}$.) and $1,100^\circ \text{C}$. ($2,012^\circ \text{F}$.).

Dr Otto Dormann (*Ton.-Industrie Zeitung*, 1914, 38, 150, 405, 1741) states that carbon dioxide begins to escape in traces at 400°C . (732°F .). It is perceptibly rapid at 700°C . ($1,292^\circ \text{F}$.), and at 800° to 900°C . ($1,472$ to $1,652^\circ \text{F}$.) for the most part has been expelled.

The cement raw material, however, only loses the last trace of carbon dioxide at $1,100^\circ \text{C}$. ($2,012^\circ \text{F}$.).

Dr A. Møye ‡ (1920) states that the temperature at which limestone begins to lose its carbon dioxide in a lime kiln depends to some extent upon the pressure of the surrounding pressure of the CO_2 . Consequently, the evolution of CO_2 does not begin at, say, 500°C . (932°F .), as it would in the absence of a surrounding atmosphere of CO_2 , but first at 800°C . ($1,472^\circ \text{F}$.). At 900°C . ($1,652^\circ \text{F}$.) the dissociation pressure reaches the average pressure of the atmosphere, viz., 15 lbs. per sq. in., and consequently the free evolution of the gas is not hindered after this temperature is reached by the pressure of the surrounding carbon dioxide.

W. J. Cooper § (1921) states that in his own experience he found calcium carbonate to be decomposed in part below 600°C . ($1,112^\circ \text{F}$.) and to a great extent at or about 700°C . ($1,292^\circ \text{F}$.), but that the time factor appears to be of considerable importance.

* *Grunberg*, *Zeitsch. Anorg. Chem.*, 1913, 80, 337.

† Dissertation, *Tech. Hochschule, Dresden*, 1912; also *Zeits. f. Elektrochem.*, 1912, 18, 729.

‡ *A. Møye*, "Theorie des Kalkbrennens," *Ton.-Industrie Zeitung*, 1920, 44, 524, Nr. 54.

§ "Consumption of Heat in Rotary Kilns," "British Portland Cement Research Association Quarterly Report for Three Months ending 31st December 1921," p. 9.

The latter author gives the following additional references to —

Eckel, $1,300^{\circ}\text{F} = 708^{\circ}\text{C}$

Meade, $900^{\circ}\text{F} = 482^{\circ}\text{C}$, which seems a low figure

Blount, $812^{\circ}\text{C} = 1,494^{\circ}\text{F}$ He further states that dissociation begins at about $400^{\circ}\text{C} = 752^{\circ}\text{F}$

C E Davis, $2,000^{\circ}\text{F} = 1,093^{\circ}\text{C}$

Martin, $812^{\circ}\text{C} = 1,494^{\circ}\text{F}$ at 753 mm pressure

Le Chatelier, $812^{\circ}\text{C} = 1,494^{\circ}\text{F}$. Elsewhere he states that it ranges from 890°C ($1,634^{\circ}\text{F}$) to 930°C ($1,706^{\circ}\text{F}$) according to physical conditions.

Tables Annuelles, $900^{\circ}\text{C} = 1,652^{\circ}\text{F}$

Bornstein's Tabellen, 887°C ($1,629^{\circ}\text{F}$) to 911°C ($1,672^{\circ}\text{F}$)

Nernst's Curve, about $830^{\circ}\text{C} = 1,526^{\circ}\text{F}$

Journal American Chemical Society, p 938, August 1910 A carefully performed estimation gives the figure as $898^{\circ}\text{C} = 1,648^{\circ}\text{F}$

† *Pierre Jolibois et Bouvier, Comptes Rendus*, 1921, 172, 1182-1183.

From (1)

$$Q = -RT^2 \times \frac{d \log p}{dT},$$

or

$$Q = 42700 + 2.183T - 0.005485T^2 \quad (3)$$

From this formula we are enabled to calculate the heat of dissociation of calcium carbonate at any temperature. For example, we have proved that in the calcining zone of a rotary kiln the initial temperature of decomposition of the calcium carbonate is $805^\circ \text{C} = 805 + 273 = 1,078^\circ \text{C}$ absolute, whence $T = 1078$, substituting in (3)

$$\begin{aligned} Q_{805^\circ \text{C}} &= 42700 + 2.183 \times 1078 - 0.005485(1078)^2 \\ &= 38,680 \text{ gram calories per 100 grams of calcium carbonate decomposed} \\ &= 696.3 \text{ B Th U's per lb } \text{CaCO}_3 \text{ decomposed} \end{aligned}$$

Farther up the calcining zone nearer the sintering zone, we proved that the initial temperature of decomposition is $784^\circ \text{C} = 1,057^\circ$ absolute

$$\begin{aligned} \text{Hence } Q_{784^\circ \text{C}} &= 42700 + 2.183 \times 1057 + 0.005485(1057)^2 \\ &= 38,879 \text{ gram calories per 100 grams } \text{CaCO}_3 \text{ decomposed} \\ &= 699.8 \text{ B Th U's per lb of } \text{CaCO}_3 \end{aligned}$$

$$\text{Similarly, } Q_{850^\circ \text{C}} = 688 \text{ B Th U's decomposed}$$

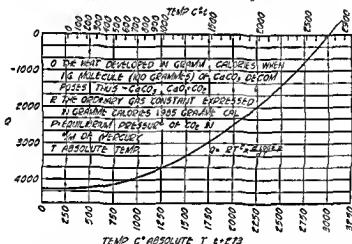


FIG 1—Curve Showing Variation with Temperature in the Molecular Heat of Dissociation of Calcium Carbonate Expressed in Gram Calories per 100 Grams of Calcium Carbonate

Proceeding in this manner, we have calculated in Table I (columns C and E) the heat of dissociation between -73°C and $+3,000^\circ \text{C}$, and the results are plotted in the two curves—Fig 1 which gives the molecular heat of dissociation, i.e. the heat required to decompose 100 grams of calcium carbonate, and Fig 2, which gives the same result in B Th U's per lb of CaCO_3 and in Fahrenheit degrees

The results shown in columns C and E are calculated from J. Johnstone's results (see *Jour Amer Chem Soc*, 1910 32, p 943) from the dissociation pressures of the evolved carbon dioxide, who in his turn assumed that the heat of dissociation of calcium carbonate at 27°C per gram molecule (100 grams) is

$$Q_{27} = 42,900 \text{ gram-calories}$$

This was the mean value of Thomsen and de Forcrand's results for precipitated chalk (42,500 and 43,300 gram calories respectively), which Johnstone took as the best value of this thermal constant (see "Monthly Report" for September 1911, p 18)

TABLE I.—Heat of Dissociation of Calcium Carbonate (CaCO_3) at Different Temperatures

| Temperature Degrees. | | Heat of Dissociation for 1 Gram-Molecule (100 Grams) of CaCO_3 . | | B.Th. U.'s Required to Decompose 1 Lb. of CaCO_3 . | | Remarks. |
|----------------------|--------|--|--|--|--|---|
| A. | B. | For Precipitated Chalk Based on J. Johnstone's Value 42,900 Gram-Calories at 27° C.* | For Calcite Based on de Forcrand's Value 42,000 Gram-Calories at 27° C.† | For Precipitated Chalk Based on J. Johnstone's Value 42,900 Gram-Calories at 27° C.* | For Calcite Based on de Forcrand's Value 42,000 Gram-Calories at 27° C.† | |
| ° C. | | Gram-Calories. | | B.Th. U.'s. | | |
| ° F. | | Gram-Calories. | | B.Th. U.'s. | | |
| — 73° | 109.4° | —42,917 | —42,015 | —772.6 | —756.40 | |
| + 27° | 80.6° | —42,900 | —42,000 | —772.2 | —755.98 | |
| + 100° | 212° | —42,751 | —41,853 | —769.5 | —753.3 | |
| + 200° | 392° | —42,506 | —41,613 | —765.2 | —749.1 | |
| + 300° | 572° | —42,151 | —41,266 | —758.6 | —742.6 | |
| + 400° | 752° | —41,687 | —40,811 | —750.4 | —734.6 | |
| + 500° | 932° | —41,113 | —40,249 | —740.0 | —724.5 | |
| + 600° | 1,112° | —40,429 | —39,580 | —729.8 | —714.5 | |
| + 700° | 1,292° | —39,636 | —38,803 | —713.5 | —698.5 | |
| + 784° | 1,443° | —38,879 | —38,062 | —699.8 | —685.1 | |
| + 800° | 1,472° | —38,733 | —37,919 | —697.2 | —682.5 | Lower limit of initial decomposition in rotary kiln. |
| + 805° | 1,481° | —38,680 | —37,868 | —696.3 | —681.6 | Upper limit of initial decomposition in rotary kiln. |
| + 900° | 1,652° | —37,720 | —36,928 | —679.0 | —664.7 | |
| + 1,000° | 1,832° | —36,598 | —35,829 | —658.0 | —644.2 | |
| + 2,000° | 3,632° | —19,349 | —18,942 | —348.4 | —341.1 | |
| + 2,727° | 4,940° | + 71 | + 69.5 | + 1.88 | + 1.25 | At this temperature practically no heat is required to decompose the CaCO_3 . Notice change from — to +. |
| + 3,000° | 5,432° | + 859.7 | + 8,673 | + 159.5 | + 156.1 | |

* J. Johnstone, Jour. Amer. Chem. Soc., 1910, 32, p. 941.

† De Forcrand, Comptes Rendus, 1908, 146, p. 512.

‡ In the cement industry nearly all the raw material is in the form of *calcite*. Hence the results of columns D and F are the values of practical importance.

However, we know that in the raw materials employed in the manufacture of Portland cement the calcium carbonate is mainly in the form of *calcite*, and hence a better value to adopt, for the heat of formation of calcium carbonate is the value for *calcite*, viz., $Q_{27} = 42,000$ gram calories per 100 grams CaCO_3 , thus being the value arrived at by de Forcrand (*Comptes Rendus*, 1908, 146, p. 512) for this material.

For the cement industry, therefore, we should correct the values shown in columns C and E by multiplying by the factors $\frac{42000}{42900}$, or $\frac{756}{772.2}$ ($= 0.979$).

We thus arrive at the results shown in columns D and F.

Extrapolation of Results

The formula

$$Q = 42700 + 2.183T + 0.005485T^2 \quad (3)$$

has been derived from experimental data obtained at temperatures ranging from 18° to 894°C , and consequently can only be considered as correct between these two temperature limits.

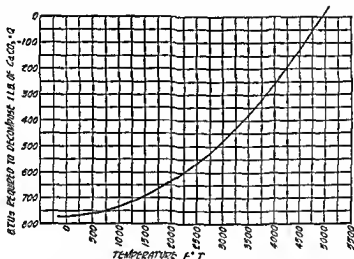


FIG. 2.—Curve Showing Variation with Temperature in the Heat Required to Decompose Calcium Carbonate Expressed in B Th U's per 1 lb. of Calcium Carbonate.

Nevertheless, the results obtained by extrapolation outside those limits are interesting, in order to see how the heat of decomposition of calcium carbonate alters with the temperature.

From the curves (and using Johnstone's value of $Q_{27} = 42,900$ gram calories) it will be seen that at ordinary temperatures (27°C) the heat of decomposition is -772.6 B Th U's (i.e., heat is *absorbed* in decomposing the CaCO_3 into CaO and CO_2), but that this quantity steadily becomes *less* as the temperature increases, until at about $2,726^\circ \text{C}$ the value is *zero*. At temperatures *above* that, the heat of formation becomes *+*ve, i.e., the compound evolves heat on decomposition—which at $3,000^\circ \text{C}$ is $+159$. If capable of existing under an enormous pressure at $3,000^\circ \text{C}$, CaCO_3 would therefore *evolve* heat on decomposing and would therefore be in the same state as explosives like nitroglycerine are in at ordinary temperatures.

If we could practically heat the calcium carbonate in a confined space to about $2,726^\circ \text{C}$, we could separate from it at this temperature all the carbon dioxide without absorbing any heat.

Practically it would be impossible to do this because of the enormous pressure which would be required to keep the CaCO_3 from splitting up to CaO and CO_2 before the critical temperature of $2,726^\circ \text{C}$ was reached.

CHAPTER IX

CALCULATION OF THE QUANTITY OF HEAT
ABSORBED IN MAKING 1 LB. OF PORTLAND
CEMENT CLINKER

Calculation of the Quantity of Pure Dry Raw Material
Required to Make 100 Lbs. of Portland Cement Clinker

Assume for Portland cement clinker the following composition :—

| | | |
|---------------------------------|-----------|--------|
| Calcium oxide, CaO | | 64.00 |
| Alumina, Al_2O_3 | | 7.00 |
| Ferric oxide, Fe_2O_3 | | 3.00 |
| Silica, SiO_2 | | 22.00 |
| Other non-essential ingredients | | 4.00 |
| | | <hr/> |
| | | 100.00 |

Omitting the unessential ingredients, the clinker would reduce to the following composition :—

| | |
|-------------|-------------|
| CaO = | 66.27 lbs. |
| Al_2O_3 = | 7.29 " |
| Fe_2O_3 = | 3.12 " |
| SiO_2 = | 22.92 " |
| | <hr/> |
| | 100.00 lbs. |

We will take this *simplified clinker* as the basis for our calculations in the following pages.

The dried slurry required to produce 100 lbs. of this clinker will have the following composition :—

| | | |
|--|--------------------------|--|
| Calcium carbonate, $CaCO_3$ | . . . 119.05 lbs. | { 66.67 lbs. CaO 52.38 " CO_2 |
| Kaolin, $Al_2Si_2O_7 \cdot 2H_2O$ | . . . 18.46 " | { 7.29 " Al_2O_3 8.60 " SiO_2 |
| | | { 2.57 " H_2O |
| Hydrated silica { 93 per cent. SiO_2 } | . . . 15.40 " | { 1.08 " H_2O 14.32 " SiO_2 |
| Ferric oxide, Fe_2O_3 | . . . 3.12 " | |
| | | <hr/> |
| Total | . 156.03 lbs. containing | { 66.67 lbs. CaO 52.38 " CO_2 7.29 " Al_2O_3 22.92 " SiO_2 3.65 " H_2O 3.12 " Fe_2O_3 |

The dried slurry on ignition will lose 35.9 per cent.

Calculation of the Minimum Quantity of Heat which Must be Liberated below 805°C ($1,481^{\circ}\text{F}$) in Order to Make 100 Lbs of Clinker

In order to make 100 lbs of clinker we must start with

| | |
|-----------------------|------------|
| (a) CaCO_3 . | 119.05 lbs |
| (b) Kaolin | 18.46 " |
| (c) Hydrated silica | 15.40 " |
| (d) Ferric oxide | 3.12 " |

Assume that the initial temperature is 60°F (15.6°C .)

(a) *Calcium Carbonate*—The mean specific heat of CaCO_3 between 16° and 805°C is 0.267*

Hence the quantity of heat in *British Thermal Units* required to raise 119.05 lbs of CaCO_3 from 60°F . (15.6°C) to $1,481^{\circ}\text{F}$ (805°C) is

$$119.05 \times 0.267 \times 1421 = 45,168 \text{ B Th U's}$$

(b) *Kaolin*—The mean specific heat of kaolin between 20°C (68°F) and 98°C (208.4°F .) is given by Ulrich as 0.224† It will certainly be greater than this at higher temperatures because specific heats increase with the temperature, also the kaolin decomposes between 400° and 800°C . According to Mellor and Holdcroft,‡ the quantity of heat required to dehydrate 1 lb of kaolin is 75.5 B Th U's. After dehydrating, kaolin is supposed to be decomposed between 450°C (842°F) and 900°C ($1,652^{\circ}\text{F}$) into a mixture of alumina, Al_2O_3 , and silica, SiO_2 .

One lb of kaolin, $\text{Al}_2\text{Si}_2\text{O}_7 \cdot 2\text{H}_2\text{O}$, contains 0.1391 lb H_2O , 0.3949 lb Al_2O_3 , 0.4660 lb SiO_2 .

Hence we may proceed as follows—

(1) Calculate the amount of heat required to dehydrate 1 lb of kaolin at 60°F . Assume this to be 75.5 B Th U's per 1 lb of kaolin.

(2) After dehydrating the 1 lb of kaolin we have left 0.3949 lb of Al_2O_3 , 0.4660 lb of SiO_2 , and 0.1391 lb of water.

We must then calculate the quantity of heat required to raise these materials from 60°F (15.6°C) to $1,481^{\circ}\text{F}$ (805°C)§. The total will be the quantity of heat required to raise 1 lb of clay to 805°C ($1,481^{\circ}\text{F}$).

To make this calculation we require the mean specific heat of SiO_2 and Al_2O_3 between ordinary temperatures and 805°C ($1,481^{\circ}\text{F}$). According to Richards,|| the mean specific heat S of Al_2O_3 between 0° and $t^{\circ}\text{C}$ is given by the formula

$$S_a = 0.2081 + 0.0000876t$$

Substituting $t = 805^{\circ}$, we obtain for the mean specific heat of Al_2O_3 between 0° and 805°C

$$\begin{aligned} &= 0.2081 + 0.0000876 \times 805 \\ &= 0.2081 + 0.0705 \\ &= 0.2786 \end{aligned}$$

Hence the quantity of heat required to raise 0.3949 Al_2O_3 from 60°F (15.6°C) to $1,481^{\circ}\text{F}$ (805°C) is $0.3949 \times 0.2786 \times 1421$

$$\begin{aligned} &= 0.11001914 \times 1421 \\ &= 156.337 \text{ B Th U's} \end{aligned}$$

* By interpolation from Magnus's "Determinations of the Specific Heat of Calcium Carbonate." See *A Magnus Ann d Phys* 1910 31 (4) 597. *Phys Zeits* 1913 14 5.

† Ulrich, *Wollny Forsch a d Geb d Agriculturphys* 1894 17 1.

‡ Mellor and Holdcroft. Collected Papers from the County Pottery Laboratory, Staffordshire, vol 1 p 277 (Griffin London, 1914).

§ Clay takes about thirty minutes to dehydrate at 500°C . Only a small amount of water vapour escapes in thirty minutes at 500°C , so that the assumption that the whole of the water vapour is raised to 805°C before it escapes will not cause a serious error.

|| J W Richards "Metallurgical Calculations" 1917, part 1, p 123.

Also according to Richards (*loc. cit.*), the mean specific heat S_m of SiO_2 between 0° and t° C. is given by the formula

$$S_m = 0.1833 + 0.000077t.$$

Substituting $t = 805^\circ$ C.,

$$\begin{aligned} S_m &= 0.1833 + 0.0620 \\ &= 0.2453. \end{aligned}$$

Hence the quantity of heat required to raise 0.4660 lb. of SiO_2 from 60° F. (15.5° C.) to $1,481^\circ$ F. (805° C.) is $0.4660 \times 0.2453 \times 1421 = 162.434$ B.Th.U.'s.

As regards the combined water (0.1391 lb.) in the kaolin, we assume that this is (1) separated in liquid form from the kaolin at 60° F. (15.6° C.); (2) heated to 212° F.; (3) turned into vapour at 212° F.; and (4) the temperature of the vapour increased from 212° F. (100° C.) to $1,481^\circ$ F. (805° C.).

(1) To separate 0.1391 lb. of water from the kaolin at 60° F. (15.6° C.) is assumed to absorb 75.5 B.Th.U.'s.

(2) To raise 0.1391 lb. of water from 60° to 212° F. requires

$$0.1391 \times 152 = 21.1 \text{ B.Th.U.'s.}$$

(3) To gasify 0.1391 lb. of water at 212° F. under atmospheric pressure requires

$$0.1391 \times 970.7 = 135.0 \text{ B.Th.U.'s.}$$

(4) From the B.P.C.R. Association tables the amount of heat in B.Th.U.'s required to raise 1 lb. of water vapour from 212° F. (100° C.) to $1,481^\circ$ F. (805° C.) is

$$696.2 - 84.2 = 612.0 \text{ B.Th.U.'s.}$$

Hence the quantity of heat to raise 0.1391 lb. through the same range of temperature is

$$0.1391 \times 612 = 85.1 \text{ B.Th.U.'s.}$$

Hence the total quantity of heat required to dehydrate 1 lb. of kaolin and raise it from 60° F. (15.6° C.) to $1,481^\circ$ F. (805° C.) is

$$156.3 + 162.4 + 75.5 + 21.1 + 85.1 + 135.0 = 635.4 \text{ B.Th.U.'s.}$$

But to make 100 lbs. of clinker it is necessary to heat 18.46 lbs. of kaolin from 60° F. (15.5° C.) to $1,481^\circ$ F. (805° C.) in order to bring it to the temperature at which the CaCO_3 decomposes.

Hence the total quantity of heat required for this operation is

$$18.46 \times 635.4 = 11,729 \text{ B.Th.U.'s.}$$

(c) *Hydrated Silica*.—There are 15.40 lbs. of hydrated silica (consisting of 14.32 lbs. of SiO_2 united to 1.08 lbs. of water), which is to be heated from 60° F. (15.6° C.) to $1,481^\circ$ F. (805° C.).

Here again we will proceed by assuming:—

(1) That the silica is dehydrated at 60° F. (15.6° C.).

(2) That the resulting water is then heated from 60° F. (15.6° C.) to 212° F. (100° C.).

(3) The water is turned into vapour at 212° F. (100° C.).

(4) The vapour is then heated from 212° F. (100° C.) to 752° F. (400° C.).

(5) The silica is heated from 60° F. (15.6° C.) to $1,481^\circ$ F. (805° C.), the latter being the temperature at which the silica is completely dehydrated.

(1) The literature and known data relating to the dehydration of silica acids is given in the "Progress Report of the British Portland Cement Research Association for the Three Months ending 31st March 1923," p. 57.

The quantity of heat required to dehydrate 1 lb. of the hydrated silica is unknown. Seeing, however, that the dehydration is complete at 400° C. against

800° C for kaolin, we will assume, in order to obtain a working figure, that the quantity of heat required for this purpose is only one third of that required to dehydrate an equal weight of clay

To dehydrate 1 lb of kaolin requires 75.5 B Th U's. Hence, to dehydrate 1 lb of our hydrated silica will require 25.2 B Th U's

(2) The amount of heat required to heat the separated water in 1 lb of the hydrated silica (0.07 lb) from 60° to 212° F is

$$0.07 \times 152 = 10.6 \text{ B Th U's}$$

(3) Amount of heat required to change 0.07 lb of water into steam at 212° F and atmospheric pressure is

$$0.07 \times 970.7 = 79.5 \text{ B Th U's}$$

(4) The amount of heat required to raise 1.00 lb of steam from 212° F (100° C) to 752° F (400° C) is from the B P C R Association tables —

$$336.3 - 84.3 = 252 \text{ B Th U's}$$

Hence the amount required for 0.07 lb is

$$252 \times 0.07 = 17.64 \text{ B Th U's}$$

(5) The amount of heat required to raise 0.93 lb of silica, SiO_2 , from 60° F (15.6° C) to 1,481° F (805° C) is

$$0.93 \times 0.2453 \times 1421 = 324.2 \text{ B Th U's}$$

Hence the total amount of heat required to dehydrate 1 lb of hydrated silica and raise its temperature from 60° F (15.6° C) to 1,481° F (805° C) is

$$75.2 + 10.6 + 79.5 + 17.6 + 324.2 \\ = 457.1 \text{ B Th U's}$$

But to make 100 lbs of clinker we must treat in this manner 15.40 lbs of hydrated silica

Hence the total amount of heat required for this operation is

$$15.40 \times 457.1 = 7,039 \text{ B Th U's}$$

(d) *Ferric Oxide, Fe_2O_3* —In making 100 lbs of clinker, 3.12 lbs of ferric oxide, Fe_2O_3 , are to be heated from 60° F (15.6° C) to 1,481° F (805° C)

According to J W Richards ("Metallurgical Calculations," 1917, part 1, p 125), the mean specific heat of ferric oxide, Fe_2O_3 , between 0° and t° C is given by the formula

$$S_m = 0.1456 + 0.000188t$$

Substituting $t = 805^\circ \text{C}$, we get for the mean specific heat between 0° and 805° C

$$S_m = 0.1456 + 0.000188 \times 805 \\ = 0.1579$$

Hence the quantity of heat required to raise 3.12 lbs of ferric oxide from 60° to 1,481° F is

$$3.12 \times 0.1579 \times 1421 \\ = 0.92633 \times 1421 \\ = 1,316 \text{ B Th U's}$$

Summary—The quantity of heat which must be expended below 805° C (1,481° F) in order to make 100 lbs of clinker is the sum of (a), (b), (c), and (d) —

$$\begin{array}{cccc} = & (a) & (b) & (c) & (d) \\ = & 45168 & + & 117.9 & + & 7039 & + & 1316 \\ = & 65,252 & \text{B Th U's,} \end{array}$$

or 1 lb of clinker requires to be expended below 805° C (1,481° F)

$$652.5 \text{ B Th U's}$$

Calculation of the Minimum Quantity of Heat which Must be Liberated above 805° C. (1,481° F.) in Order to Make 100 Lbs. of Clinker

We assume, as above, that 100 lbs. of clinker are produced from a dry slurry containing:—

| | |
|-------------|--|
| 119.05 lbs. | pure calcium carbonate, CaCO_3 . |
| 18.46 " | kaolin, $\text{Al}_2\text{Si}_2\text{O}_7\cdot 2\text{H}_2\text{O}$. |
| 15.40 " | hydrated silica (93 per cent. SiO_2 ; 7 per cent. H_2O). |
| 3.12 " | ferric oxide. |

Total, 156.03 lbs.

The calcium carbonate begins to decompose in the furnace with evolution of CO_2 at 805° C. (1,481° F.) at this temperature, and before the calcium carbonate begins to decompose, we may assume that all the kaolin and silica have lost. Their combined water and the kaolin have been resolved into a mass of alumina and silica, so that the mass will consist of

| | | | | | | |
|---------------------------------------|---|---|---|-------------|--------------|--------------------------------|
| Calcium carbonate, CaCO_3 | . | . | . | 119.05 lbs. | { | 66.67 lbs. CaO |
| | | | | | | 52.38 " CO_2 |
| Dehydrated kaolin | . | . | . | 15.89 " | { | 7.29 " Al_2O_3 |
| | | | | | | 8.60 " SiO_2 |
| Dehydrated silica, SiO_2 | . | . | . | . | | 14.32 " SiO_2 |
| Ferric oxide, Fe_2O_3 | . | . | . | . | | 3.12 " |
| | | | | | } 22.92 lbs. | |

Total weight . 152.38 lbs.

The 119.05 lbs. of CaCO_3 when decomposed yield 52.38 lbs. of CO_2 and 66.67 lbs. of CaO .

So that as soon as the expulsion of the carbon dioxide is completed, the solid mass will consist of

| | | | | | |
|---------------------------------------|---|---|---|---|------------|
| Lime, CaO | . | . | . | . | 66.67 lbs. |
| Alumina, Al_2O_3 | . | . | . | . | 7.29 " |
| Silica, SiO_2 | . | . | . | . | 22.92 " |
| Ferric oxide, Fe_2O_3 | . | . | . | . | 3.12 " |

100.00 lbs.

This mass will unite to form clinker with the evolution of some heat.

The necessary calculation may be carried out as follows:—

(a) Calculate the amount of heat absorbed in decomposing 119.05 lbs. of CaCO_3 at 805° C. (1,481° F.).

It has been shown that the decomposition of 1 lb. of CaCO_3 at this temperature absorbs 682 B.Th.U.'s.

Hence, to decompose 119.05 lbs. of CaCO_3 requires

$$119.05 \times 682 = 81,192 \text{ B.Th.U.'s.}$$

(b) Calculate the amount of heat evolved in the exothermic reaction when 66.67 lbs. of CaO , 7.29 lbs. of Al_2O_3 , 22.92 lbs. SiO_2 , and 3.12 lbs. of Fe_2O_3 unite to form 100 lbs. of Portland cement clinker at 805° C. (1,481° F.).

The recent extremely accurate results of R. Nacken in 1922 (see Pamphlet No. 3 of the B.P.C.R.A., "The Exothermic Reaction of Portland Cement Clinker Formation," Table 14, p. 14) lead to the result that, to every 1 lb. of clinker produced from the slurry, 179.93 B.Th.U.'s are evolved.

So that when 100 lbs. of clinker are formed, there are evolved 17,993 B.Th.U.'s.

(c) Calculate the amount of heat absorbed in raising the resultant 100 lbs. of clinker from 805° C. (1,481° F.) to the highest clinkering temperature of, say, 1,370° C. (2,498° F.).

Very careful determinations of the specific heat of clinker at high temperatures have recently been made by *S. H. Harrison* in conjunction with Dr W. D. White of the Carnegie Institution of Washington Geophysical Laboratory ("Report on the Specific Heat of Portland Cement Clinker," published by the Committee on Conservation Portland Cement Association, 1923)

The specific heat of any unchanged lime (CaO) in the mass is taken as nearly equal to that of the clinker. *Harrison (loc cit)* gives the following formula for calculating the amount of heat expressed in centigrade units required to raise 1 lb of the clinker from 0° to $t^\circ \text{C}$ —

$$Q = 0.1754t + 139144 \times 10^{-9}t^2 - 125 \times 10^{-9}t^3 + 4685 \times 10^{-14}t^4$$

Substituting in succession $t = 1,370^\circ \text{C}$ and $t = 805^\circ \text{C}$ in this formula and subtracting, we obtain the amount of heat (in centigrade units) required to raise 1 lb of clinker from 805°C . ($1,481^\circ \text{F}$) to $1,370^\circ \text{C}$ ($2,498^\circ \text{F}$) as follows —

$$\begin{aligned} Q &= 0.1754(1370 - 805) + 139144 \times 10^{-9}(1370^2 - 805^2) - 125 \times 10^{-9}(1370^3 - 805^3) \\ &\quad + 4685 \times 10^{-14}(1370^4 - 805^4) \\ &= 159.25 \text{ centigrade heat units (1 lb water through } 1^\circ \text{C)} \\ &= 159.25 \times 1.8 \text{ B Th U's} \\ &= 286.65 \text{ B.Th U's (1 lb water through } 1^\circ \text{F)} \end{aligned}$$

The mean specific heat of the clinker between 805° and $1,370^\circ \text{C}$ ($1,481^\circ$ to $2,606^\circ \text{F}$) is

$$\frac{286.65}{565} = 0.507$$

Hence the total amount of heat required to raise 100 lbs. clinker from 805°C ($1,481^\circ \text{F}$) to $1,370^\circ \text{C}$ ($2,498^\circ \text{F}$) is

$$28,665 \text{ B.Th U's.}$$

Summary — We are now in a position to calculate the minimum quantity of heat required to be absorbed *above* 805°C ($1,481^\circ \text{F}$) in order to produce 100 lbs of clinker

This quantity of heat Q is

| | | |
|---|---|--------------------------|
| Heat absorbed in decomposing 119.05 lbs of CaCO_3 | = | - 81,192 B Th U's |
| + Heat evolved in the exothermic reaction | = | + 17,993 " |
| + Heat absorbed in raising 100 lbs of clinker from 805°C ($1,481^\circ \text{F}$) to $1,370^\circ \text{C}$ ($2,498^\circ \text{F}$) | = | - 28,665 " |
| Total | | - 91,864 B Th U's |

Hence, in order to produce 100 lbs of clinker, we must expend *above* 805°C 91,864 B Th U's, or 1 lb of clinker requires 918.6 B Th U's expended on it *above* 805°C ($1,481^\circ \text{F}$), but only 652.5 B Th U's *below* 805°C ($1,481^\circ \text{F}$) *

* It should be carefully noted that it is a fallacy to state that the total *minimum quantity* of heat required to make 1 lb of clinker is $918.6 + 652.5 = 1571.1$ B Th U's because this statement takes no account of the quantity of heat recoverable from the hot clinker and evolved gases. Allowance is made for these quantities later.

All that the above statements mean is that in making 1 lb clinker the raw material mixture must be subjected to a steadily increasing temperature, starting at 60°F and finishing at about $2,498^\circ \text{F}$, and that the raw mixture must absorb 652.5 B Th U's *below* $1,481^\circ \text{F}$ and 918.6 B Th U's *above* $1,481^\circ \text{F}$ before 1 lb clinker can be formed.

CHAPTER X

CALCULATION OF THE QUANTITY OF GASES
OF COMBUSTION AND AIR QUANTITIES RE-
QUIRED TO MAKE 1 LB. OF CLINKER

THERE is a very considerable uniformity of composition of the gaseous products of combustion evolved in the making of Portland cement clinker.

For this purpose we will compare the products of combustion evolved by burning equivalent quantities of two coals—one of 12,710 B.Th.U.'s per lb. and the other of 11,851 B.Th.U.'s per lb.

Products of Combustion from a Coal of 12,710 B.Th.U.'s per 1 Lb. Compared to Products of Combustion from Standard Coal

The coal in the Works Test, No. 26, had the following composition :—

| | Dried. | As Fired. |
|---------------------|---------------|---------------|
| C | 72.202 | 71.985 |
| H | 4.99 | 4.975 |
| S | 1.18 | 1.177 |
| Ash | 17.227 | 17.175 |
| Moisture | 0.000 | 0.300 |
| N | 1.180 | 1.177 |
| N | 1.428 | 1.424 |
| O (diff.) | 2.973 | 2.964 |
| | <hr/> 100.000 | <hr/> 100.000 |

One lb. of dry coal yields when fired

| | |
|--------------------------------------|-------------------|
| CO ₂ | 2.647 lbs. |
| N | 7.736 „ |
| H ₂ O | 0.452 lb. |
| | <hr/> 10.835 lbs. |
| Add 5 per cent. excess air | 0.542 lb. |
| | <hr/> 11.377 lbs. |

One lb. of standard coal of 12,600 B.Th.U.'s per lb. is equivalent to $\frac{12600}{12710} = 0.9913$ lb. of this coal. 0.9913 lb. of this coal, therefore, yields 12,600 B.Th.U.'s, and will yield

| | |
|----------------------------|------------------------------------|
| CO ₂ | $0.9913 \times 2.647 = 2.624$ lbs. |
| N | $0.9913 \times 7.736 = 7.669$ „ |
| H ₂ O | $0.9913 \times 0.452 = 0.448$ lb. |
| Excess air | $0.9913 \times 0.542 = 0.537$ „ |
| | <hr/> 11.278 lbs. of gas. |

The percentage composition of this mixture by weight is

| | |
|------------------|-------------|
| CO ₂ | 23 267 lbs |
| N ₂ | 67 996 " |
| H ₂ O | 3 973 " |
| Excess air | 4 764 " |
| | <hr/> |
| | 100 000 lbs |

Products of Combustion from a Coal of 11,851 B Th U's per 1 Lb Compared to Products from Standard Coal

This coal had the following composition when dried —

| | | | Mean |
|-----|--------|--------|--------|
| C | 65 98 | 65 74 | 65 86 |
| H | 4 55 | 4 55 | 4 55 |
| S | 1 32 | 1 42 | 1 37 |
| Ash | 19 76 | 19 76 | 19 76 |
| O | 7 00 | 7 13 | 7 06 |
| N | 1 39 | 1 40 | 1 40 |
| | <hr/> | <hr/> | <hr/> |
| | 100 00 | 100 00 | 100 00 |

One lb dry coal yields when burnt

| | |
|---------------------------|------------|
| CO ₂ | 2 41 lbs |
| N ₂ | 6 84 " |
| H ₂ O | 0 41 lb |
| | <hr/> |
| | 9 660 lbs |
| Add 5 per cent excess air | 0 483 lb |
| | <hr/> |
| | 10 143 lbs |

One lb of standard coal of 12,600 B Th U's per lb is equivalent to $\frac{12600}{11851} = 1.063$ lbs of this coal. 1.063 lbs of this coal, therefore, yield 12,600 B Th U's, and will yield

| | |
|------------------|-----------------------------------|
| CO ₂ | $2.41 \times 1.063 = 2.56183$ lbs |
| N ₂ | $6.84 \times 1.063 = 7.27092$ " |
| H ₂ O | $0.41 \times 1.063 = 0.43583$ lb |
| Excess air | $0.483 \times 1.063 = 0.51343$ " |

10 78201 lbs of gas

corresponding to 12,600 B Th U's liberated

The percentage composition of the mixture is by weight

| | |
|------------------|-------------|
| CO ₂ | 23 760 lbs |
| N ₂ | 67 436 " |
| H ₂ O | 4 042 " |
| Excess air | 4 762 " |
| | <hr/> |
| | 100 000 lbs |

Calculation of the Quantity of Air Required to Produce Weights of Combustion Gas Equivalent in Heating Power

From the preceding calculations it is seen that the liberation of (a) 11 278 lbs of gas produced from the combustion of a coal of 12,710 B Th U's per lb, and (b) 10 782 lbs of gas produced from the combustion of a coal of 11,851 B Th U's per lb each correspond to the liberation in the furnace of 12,600 B Th U's

from the coal—i.e., are equivalent to the combustion of 1 lb. of standard coal in the furnace. We wish now to calculate how much *air* has to be supplied (a) to make the 11.278 lbs. of gas, and (b) to make the 10.782 lbs. of gas in the two cases above mentioned. For obviously, if the cold air is drawn into the furnace from the atmosphere, the heat must be supplied for preheating this cold air, and if the quantities of air so supplied are different, obviously different quantities of heat must be supplied for this purpose. The matter is, therefore, of considerable practical importance.

In case (a) 0.9913 lb. of coal has to be burnt to liberate 12,600 B.Th.U.'s.

In case (b) 1.063 lbs. of coal have to be burnt to liberate 12,600 B.Th.U.'s.

Coal (a).—Calorific value, 12,710 B.Th.U.'s per 1 lb. composition.

| | | | | | | |
|-----------|---|---|---|---|---|------------------|
| C | . | . | . | . | . | 72.202 |
| S | . | . | . | . | . | 1.180 = 0.4425 C |
| H | . | . | . | . | . | 4.990 |
| Ash | . | . | . | . | . | 17.227 |
| N | . | . | . | . | . | 1.428 |
| O (diff.) | . | . | . | . | . | 2.973 |
| | | | | | | <hr/> |
| | | | | | | 100.000 |

In order to calculate the amount of air required for combustion, we assume as correct the available percentage of hydrogen as $H - \frac{O}{8}$.

$$H = 4.990 - \frac{2.973}{8.000} = 4.990 - 0.372.$$

$$H = 4.618 \text{ per cent.}$$

Hence 1 lb. of H requires 34.782 lbs. of air to burn to H_2O .

∴ 0.04618 of H requires $0.4618 \times 34.782 = 1.6062$ lbs. air.

The percentage of carbon and its equivalent of sulphur ($C = 72.202$; $S = 1.180 = 0.4425 C$.) in the coal amounts to 72.6445 per cent.

One lb. of carbon requires 11.594 lbs. of air to burn to CO_2 .

0.72645 lb. of carbon requires $11.594 \times 0.72645 = 8.4225$ lbs./air.

Hence the total amount of air required to burn 1 lb. of fine coal is

$$= 8.4225 \text{ lbs.}$$

$$1.6062 \text{ ,,}$$

$$\hline 10.0287 \text{ lbs.}$$

Weight of air needed to burn 0.9913 lb. is 10.0287×0.9913

$$= 9.94145 \text{ lbs.}$$

Add excess air

$$= 0.537 \text{ lb.}$$

$$\hline 10.478 \text{ lbs.}$$

Total air required to burn 0.9913 lb. of coal and liberate 12,600 B.Th.U.'s is 10.478 lbs. air.

Coal (b).—Calorific value, 11,851 B.Th.U.'s per 1 lb.

This coal, when dried, had the mean composition:—

| | | | | | | |
|-----|---|---|---|---|---|--------|
| C | . | . | . | . | . | 65.86 |
| H | . | . | . | . | . | 4.55 |
| S | . | . | . | . | . | 1.37 |
| Ash | . | . | . | . | . | 19.76 |
| O | . | . | . | . | . | 7.06 |
| N | . | . | . | . | . | 1.40 |
| | | | | | | <hr/> |
| | | | | | | 100.00 |

Available H for combustion is

$$4.55 - \frac{7.06}{8} = 4.55 - 0.88 = 3.67 \text{ per cent}$$

Available C for combustion + equivalent of S is

$$\begin{aligned} & 65.86 + \frac{1.37 \times 12}{32} \\ &= 65.86 + 0.51 \\ &= 66.37 \text{ per cent C} \end{aligned}$$

Hence 1 lb of this coal contains

0.0367 lb of H available for combustion
0.6637 lb of C available for combustion

Since 1 lb of H requires 34.782 lbs of air for combustion

0.0367 lb of H requires of air $0.0367 \times 34.782 = 1.2765$ lbs of air

Since 1 lb of C requires 11.594 lbs of air to burn to CO_2

0.6637 lb of C requires $11.594 \times 0.6637 = 7.6949$ lbs of air

Hence per 1 lb coal the total air required for combustion is

$$1.2765 + 7.6949 = 8.9714$$

Hence the amount of air required to burn 1.063 lbs of this coal so as to liberate 12 600 B Th U s in the kiln is

$$1.063 \times 8.9714 = 9.536 \text{ lbs of air}$$

Add to this 0.51343 lb of excess air and we get the total amount of cold air required to liberate 12 600 B Th U s —

$$9.53660 + 0.51343 = 10.05 \text{ lbs of air}$$

Summary — For the purpose of the succeeding calculations we will take it that the combustion of 1 lb of standard coal of 12 600 B Th U s per lb will yield 11.278 lbs of combustion gas composed as follows —

| <i>Gas</i> | |
|----------------------|------------|
| CO_2 | 2.624 lbs |
| N_2 | 7.669 |
| H_2O | 0.448 lb |
| Excess air | 0.537 |
| | <hr/> |
| | 11.278 lbs |

The percentage composition of this mixture by weight is

| | |
|----------------------|-------------|
| CO_2 | 23.267 lbs |
| N_2 | 67.996 |
| H_2O | 3.973 |
| Excess air | 4.764 |
| | <hr/> |
| | 100.000 lbs |

The weight of air necessary for the combustion of 1 lb of the coal so as to give a combustion gas of the preceding composition is 10.478 lbs of air

CHAPTER XI

THE SPECIFIC HEATS OF THE GASES OF COMBUSTION AT HIGH TEMPERATURES

THE main heating agent in the upper part of the kiln is the hot furnace gases evolved by the combustion of the coal dust; these gases, as previously calculated in Chapter VI., have the approximate composition by weight:—

$$\text{CO}_2 = 23.267; \text{N}_2 = 67.996; \text{H}_2\text{O} = 3.973; \text{excess air} = 4.764.$$

It is of vital importance for the theory of the rotary kiln to ascertain the specific heat of this mixture at various temperatures so as to be able to ascertain how much heat can be obtained from a given weight of this gas when it sinks through different temperature intervals as it passes down the kiln.

The author has worked out tables of the values of the *instantaneous* and *mean specific heats* at high and low temperatures of the individual gases which occur in the cement rotary kiln, these values being calculated from recent modern researches, and, consequently, they exceed in accuracy any data previously available in the cement industry.

Table I. (p. 11.2) shows these values.

The weight of *combustion gas* produced by the burning of 1 lb. of standard coal was proved in Chapter II. to be 11.278 lbs., composed of 2.624 lbs. of CO_2 ; 7.669 lbs. of N_2 ; 0.448 lb. of H_2O ; and 0.537 lb. of excess air.

The instantaneous or true specific heat S of this gaseous mixture may be calculated at the various temperatures from the formula

$$S_c = \frac{2.624 \times S_1 + 7.669 \times S_2 + 0.448 \times S_3 + 0.537 \times S_4}{11.278} \quad (1)$$

where S_1, S_2, S_3, S_4 , are the respective instantaneous specific heats of the $\text{CO}_2, \text{N}_2, \text{H}_2\text{O}$, or air at the temperature at which it is desired to ascertain the instantaneous specific heat of the mixture.

For example, the instantaneous *specific heat of the combustion gas at 4,000° F.* is obtained by substituting in the above formula the values of the *instantaneous specific heats* of the various component gases at 4,000° F., as shown in Table I. (p. 11.2).

We thus get

$$\begin{aligned} & \text{Instantaneous specific heat of combustion gas at 4,000° F.} \\ &= \frac{2.624 \times 0.32 + 7.669 \times 0.32 + 0.448 \times 0.98 + 0.537 \times 0.31}{11.276} \\ &= 0.3457. \end{aligned}$$

Proceeding in this way, the values of the *instantaneous specific heat* of the combustion gas, shown in column 1 of Table II. (p. 11.3), were obtained partly by calculation and partly by interpolation from a large-scale curve drawn through various calculated points.

A temperature of pivotal importance in the theory of the rotary kiln is 1,481° F. (805° C.), because this is the temperature at which the calcium carbonate begins to decompose under the prevailing composition of the gaseous atmosphere inside the kiln arising from the combustion of the coal.

TABLE I
Instantaneous Specific Heats of the Individual Gaseous Components of Combustion Gases

[illegible]

As will be seen in the next chapter, it is of great importance in the theory of the rotary kiln to know the mean *specific heat* of the combustion gases *between* $1,481^{\circ}$ F. (805° C.) and various temperatures both above and below $1,481^{\circ}$ F. This is calculated as follows:—

From the formula given above, the *instantaneous specific heat* of the combustion gas at $1,481^{\circ}$ F. is calculated to be 0.2799. The instantaneous specific heat of the combustion gases at any *given temperature* T° is likewise calculated to be, say, S_T . Then a *straight line law* is assumed to connect these two values (which is *nearly true*), and the formula for the *mean specific heat* S between $1,481^{\circ}$ and T° F. is

$$S = \frac{0.2799 + S_T}{2} \quad (2)$$

Proceeding in this way, column 3 of Table II. was calculated, which gives the mean specific heats up to $5,000^{\circ}$ F., on the assumption, of course, that the furnace gases remain undecomposed at these temperatures.

TABLE II

Specific Heats of Combustion Gas of a Cement Rotary Kiln at Various Temperatures°

Composition of combustion gas | $\text{CO}_2 = 23.267$ per cent.; $\text{N}_2 = 67.996$ per cent.
by weight | $\text{H}_2\text{O} = 3.973$ per cent.; excess air = 4.764.

One lb. of *standard coal* of 12,600 B.Th.U.'s per lb. evolves 11.278 lbs. of this gas consisting of $\text{CO}_2 = 2.624$ lbs.; $\text{N}_2 = 7.669$ lbs.; $\text{H}_2\text{O} = 0.448$ lb.; excess air = 0.537 lb.

| Temperature ° T. | | Instantaneous Specific Heat. | Mean Specific Heat between $1,481^{\circ}$ F. (805° C.) and T° . |
|------------------|--------|------------------------------|---|
| ° F. | ° C. | | |
| 5000 | 2760 | 0.3774 | 0.3286 |
| 4900 | 2704 | 0.3742 | 0.3270 |
| 4800 | 2649 | 0.3708 | 0.3253 |
| 4700 | 2593 | 0.3676 | 0.3237 |
| 4600 | 2538 | 0.3642 | 0.3220 |
| 4500 | 2482 | 0.3610 | 0.3204 |
| 4400 | 2427 | 0.3580 | 0.3189 |
| 4300 | 2371 | 0.3548 | 0.3173 |
| 4200 | 2316 | 0.3518 | 0.3158 |
| 4100 | 2260 | 0.3486 | 0.3142 |
| 4000 | 2204 | 0.3457 | 0.3128 |
| 3900 | 2149 | 0.3426 | 0.3112 |
| 3800 | 2093 | 0.3398 | 0.3098 |
| 3735 | 2057 | 0.3379 | 0.3089 |
| 3700 | 2038 | 0.3369 | 0.3084 |
| 3600 | 1982 | 0.3340 | 0.3069 |
| 3500 | 1926.5 | 0.3310 | 0.3054 |
| 3400 | 1872 | 0.3283 | 0.3041 |
| 3300 | 1816 | 0.3256 | 0.3027 |
| 3200 | 1760 | 0.3228 | 0.3013 |
| 3100 | 1704 | 0.3201 | 0.3000 |
| 3000 | 1649 | 0.3174 | 0.2986 |
| 2900 | 1594 | 0.3149 | 0.2974 |
| 2800 | 1538 | 0.3123 | 0.2961 |
| 2700 | 1482 | 0.3098 | 0.2948 |

TABLE II—(Continued)

| Temperature °T | | Instantaneous Specific Heat | Mean Specific Heat between 1,481° F (805° C) and T° |
|----------------|------|-----------------------------|---|
| °F | °C | | |
| 2600 | 1427 | 0 3071 | 0 2935 |
| 2590 | 1421 | 0 3068 | 0 2933 |
| 2580 | 1415 | 0 3066 | 0 2932 |
| 2570 | 1410 | 0 3063 | 0 2931 |
| 2560 | 1404 | 0 3061 | 0 2930 |
| 2550 | 1399 | 0 3058 | 0 2928 |
| 2540 | 1393 | 0 3056 | 0 2927 |
| 2530 | 1388 | 0 3053 | 0 2926 |
| 2520 | 1382 | 0 3051 | 0 2925 |
| 2510 | 1377 | 0 3048 | 0 2923 |
| 2500 | 1371 | 0 3046 | 0 2922 |
| 2490 | 1365 | 0 3044 | 0 2921 |
| 2480 | 1360 | 0 3041 | 0 2920 |
| 2470 | 1354 | 0 3039 | 0 2919 |
| 2460 | 1349 | 0 3036 | 0 2917 |
| 2450 | 1343 | 0 3034 | 0 2916 |
| 2440 | 1337 | 0 3032 | 0 2915 |
| 2430 | 1332 | 0 3029 | 0 2914 |
| 2420 | 1327 | 0 3027 | 0 2913 |
| 2410 | 1321 | 0 3024 | 0 2911 |
| 2400 | 1315 | 0 3022 | 0 2910 |
| 2300 | 1260 | 0 2998 | 0 2898 |
| 2200 | 1204 | 0 2975 | 0 2887 |
| 2100 | 1149 | 0 2952 | 0 2875 |
| 2000 | 1093 | 0 2928 | 0 2863 |
| 1900 | 1038 | 0 2903 | 0 2851 |
| 1800 | 982 | 0 2878 | 0 2838 |
| 1700 | 926 | 0 2854 | 0 2826 |
| 1600 | 871 | 0 2830 | 0 2814 |
| 1500 | 815 | 0 2804 | 0 2801 |
| 1481 | 805 | 0 2799 | 0 2799 |
| 1400 | 760 | 0 2777 | 0 2788 |
| 1300 | 704 | 0 2749 | 0 2774 |
| 1200 | 649 | 0 2720 | 0 2759 |
| 1100 | 594 | 0 2692 | 0 2745 |
| 1000 | 548 | 0 2664 | 0 2731 |
| 900 | 483 | 0 2634 | 0 2716 |
| 800 | 427 | 0 2604 | 0 2701 |
| 700 | 371 | 0 2574 | 0 2681 |
| 600 | 315 | 0 2544 | 0 2671 |
| 500 | 260 | 0 2513 | 0 2656 |
| 400 | 204 | 0 2482 | 0 2640 |
| 300 | 149 | 0 2450 | 0 2624 |
| 212 | 100 | 0 2421 | 0 2610 |
| 200 | 93 | 0 2417 | 0 2608 |
| 100 | 38 | 0 2384 | 0 2591 |
| 60 | 16 | 0 2370 | 0 2584 |
| 32 | 0 | 0 2361 | 0 2580 |

CHAPTER XII

CALCULATION OF THE THEORETICAL FLAME TEMPERATURES OF THE ROTARY KILN

Producible by the Combustion of Standard Coal with the Air Preheated to Various Amounts

To start with, we assume that 1 lb. of standard coal in burning liberates 12,600 B.Th.U.'s and forms the following gaseous mixture:—

| | | | | | | |
|------------------|---|---|---|---|---|-------------|
| CO ₂ | . | . | . | . | . | 2.624 lbs. |
| N ₂ | . | . | . | . | . | 7.669 „ |
| H ₂ O | . | . | . | . | . | 0.448 lb. |
| Excess air | . | . | . | . | . | 0.537 „ |
| | | | | | | 11.278 lbs. |

The total air (including 5 per cent. excess air) required for the combustion is put as 10.478 lbs.

We will start by assuming that the air for combustion is at 60° F. Subsequent calculations will allow for the effect of preheating the air to various amounts. The method adopted is as follows:—

We first calculate approximately the maximum temperature attained by the gas. We know that this is *about* 4,000° F. (2,204° C.). Then the *mean specific heats* of the component gases of the combustion gas are taken between 60° and 4,000° F. from the specific heat tables worked out by the writer from the most recent available data as follows:—

| | | | |
|---|---|---|--------|
| Mean specific heat of CO ₂ between 60° and 4,000° F. | . | . | 0.2771 |
| „ of N ₂ „ „ „ | . | . | 0.2769 |
| „ of H ₂ O (vapour) between 60° and 4,000° F. | . | . | 0.6261 |
| „ of air between 60° and 4,000° F. | . | . | 0.2683 |

The water is supposed to be gasified at 60° F. (15.6° C.), and then, while in a state of vapour, it is heated to its maximum temperature T° F. By interpolation from A. W. Smith's results (Glazebrook's "Dictionary of Applied Physics," 1922, vol. i., p. 554) the latent heat of evaporation of 1 gram of water at 60° F. (15.55° C.) is 587.7 gram-calories, or 1 lb. of water at 60° F. requires 587.7 × 1.8 = 1057.9 B.Th.U.'s to turn it into vapour.

We next calculate the *approximate* flame temperature T° F. attainable by the furnace gases as follows:—

$$\text{Heat liberated by 1 lb. coal} = \left\{ \begin{array}{l} \text{Heat absorbed in raising the products} \\ \text{of combustion to } T^{\circ} \text{ F.} \end{array} \right\},$$

$$\begin{aligned} \text{or} \quad 12600 &= 2.624 \times 0.2771(T - 60) + 7.669 \times 0.2769(T - 60) \\ &\quad + 0.448 \times 1057.9 + 0.448 \times 0.6261(T - 60) \\ &\quad + 0.537 \times 0.2683(T - 60), \\ \text{whence} \quad T^{\circ} &= 3,763^{\circ} \text{ F.} \end{aligned}$$

The approximation made here is the assumption that the specific heats of the gases follow a straight line law between 60° and 4000° F, which is nearly, but not quite, true.

Also the mean specific heat of the furnace gases between 60° and 4000° F is assumed as the same as that between 60° and 3763° F.

The errors thus introduced are slight, but perceptible corrections will be introduced below.

Effect of Preheated Air on the Flame Temperatures

The preceding flame temperature of 3763° F was calculated on the assumption that the entering air needed for the combustion of the coal was at the atmospheric temperature of 60° F and so did not carry in with it any heat.

If, however, the air was preheated to T° F by the time it inflamed the coal dust the quantity of heat carried by this air into the furnace is equal to the number of B Th U's required to raise the 10.478 lb. of air necessary for burning 1 lb. of standard coal from 60° to T° F.

Let the quantity of heat be denoted by Q B Th U's, then Q must be added on to the 12,600 B Th U's liberated by the coal in order to arrive at the total heat liberated in the furnace per 1 lb. of coal burnt. The presence of preheated air must therefore, raise the flame temperature.

We must now calculate a series of *approximate flame temperatures* corresponding to air preheated by various amounts, which temperatures will later be used for deducing *more exact values* in the manner described below.

For our purpose it will be sufficiently correct to assume that the mean specific heats of the gases concerned are those prevailing between 60° and 4000° F. This will result in the flame temperatures being usually slightly too high, but the error so introduced will be nearly compensated for by opposite errors in deducing the more exact results detailed below.

The former formula for calculating the flame temperature now becomes modified as follows —

$$\text{Total heat} = \left\{ \begin{array}{l} \text{Heat liberated} \\ \text{by 1 lb. of} \\ \text{coal} \end{array} \right\} + \left\{ \begin{array}{l} \text{Heat brought into} \\ \text{the furnace by} \\ \text{10.478 lb. of air} \end{array} \right\} = 12,600 \text{ B Th U's} + Q$$

$$\text{So that } 12,600 + Q = 26.4 \times 0.771(T - 60) + 7.669 \times 0.2769(T - 60) \\ + 0.418 \times 10.478 \times 0.418 \times 0.661(T - 60) \\ + 0.537 \times 0.683(T - 60)$$

$$\text{whence } T = 3763 + 0.3023Q \quad (1)$$

Q is calculated from the tables for the *radiation estimation of the heat carried away by the exit gases of cement rotary kilns* (Progress Report of the British Portland Cement Research Association for April 1924). The results are shown in Table I (p. 123).

Now, substituting the various values of Q from Table I in formula (1) we obtain the approximate *flame temperatures* corresponding to the different temperatures to which the air is preheated, shown in Table II (p. 123).

More Accurate Calculation of Flame Temperatures

The results shown in Table II are only approximate for reasons above mentioned but they can be made the basis for more accurate determinations of the flame temperatures as follows —

The furnace gases when heated from 60° F to T° (highest flame temperature degree) will absorb 12,600 B Th U's. But from the *Table* worked out by the author for the *British Portland Cement Research Association* we can find out very accurately the quantity absorbed in heating the furnace gases from 60° to $1,481^{\circ}$ F,

TABLE I

Quantity of Heat Required for Preheating 10.478 Lbs. of Air from 60° F. (15.6° C.) Various Temperatures

| Temperature° to which the Air is to be Preheated. F. | | B.Th.U.'s Required to Heat 1 Lb. of Air from 60° to F. | B.Th.U.'s Required to Heat 10.478 Lbs. of Air from 60° to F. Q. |
|---|------|--|--|
| ° F. | ° C. | | |
| 60 | 16 | 0.000 | 0.0 |
| 100 | 38 | 9.212 | 96.5 |
| 200 | 93 | 32.242 | 338.0 |
| 300 | 149 | 55.560 | 582.0 |
| 400 | 204 | 79.004 | 828.0 |
| 500 | 260 | 102.710 | 1076.0 |
| 600 | 315 | 126.609 | 1327.0 |
| 700 | 371 | 150.718 | 1579.0 |
| 800 | 427 | 175.002 | 1834.0 |
| 900 | 483 | 199.472 | 2090.0 |
| 1000 | 548 | 224.187 | 2349.0 |
| 1100 | 594 | 249.017 | 2609.0 |
| 1200 | 649 | 274.163 | 2873.0 |
| 1300 | 704 | 299.375 | 3137.0 |
| 1400 | 760 | 324.867 | 3404.0 |
| 1500 | 815 | 350.558 | 3673.0 |
| 1600 | 871 | 376.446 | 3944.0 |
| 1700 | 926 | 402.532 | 4218.0 |
| 1800 | 982 | 429.177 | 4497.0 |
| 1900 | 1038 | 455.360 | 4771.0 |
| 2000 | 1093 | 481.990 | 5050.0 |
| 2100 | 1149 | 508.937 | 5333.0 |
| 2200 | 1204 | 535.940 | 5616.0 |
| 2300 | 1260 | 563.090 | 5900.0 |
| 2400 | 1315 | 590.453 | 6187.0 |
| 2500 | 1371 | 618.363 | 6479.0 |

TABLE II

Approximate Theoretical Flame Temperature Obtainable by the Combustion Gases in a Cement Rotary Kiln

$$T = 3763 + 0.3053Q_r$$

| Temperature to which Entering Air is Preheated. | Approximate Flame Temperature. | Temperature to which Entering Air is Preheated. | Approximate Flame Temperature. |
|---|--------------------------------|---|--------------------------------|
| ° F. | T° F. | ° F. | T° F. |
| 60 | 3763 | 900 | 4401 |
| 100 | 3792 | 1000 | 4480 |
| 200 | 3865 | 1100 | 4559 |
| 300 | 3941 | 1200 | 4640 |
| 400 | 4016 | 1300 | 4721 |
| 500 | 4092 | 1400 | 4802 |
| 600 | 4168 | 1500 | 4884 |
| 700 | 4245 | 1600 | 4967 |
| 800 | 4323 | | |

and the difference between this quantity and 12,600 B Th U 's will be the amount of heat absorbed in heating the gases from 1,481° F to the final temperature T°

We will calculate these quantities

Quantity of Heat Liberated by 11 278 Lbs of Furnace Gas Sinking from 1,481° to 60° F

| Gas | B Th U 's Evolved by 1 Lb in Sinking from 1 481° to 60° F (a) | Weight of Gas (b) | Total B Th U 's (a × b) |
|---------------------------|--|----------------------|----------------------------|
| CO ₂ | 360 911 - 5 831 = 355 080 | 1 lbs | |
| N ₂ | 362 685 - 6 633 = 356 052 | 2 624 | 931 7299 |
| H ₂ O (vapour) | 696 358 - 13 107 = 683 251 | 7 669 | 2730 5628 |
| Excess air | 352 200 - 6.448 = 345 752 | 0 448 | 306 0964 |
| | | 0 537 | 185 6688 |
| | | 11 278 | 4154 0579 |

So that 11 278 lbs of furnace gas sinking from 1,481° F liberate 4154 1 B Th U 's To this must be added the quantity of heat liberated when 0 448 lb of water vapour at 60° F condenses to liquid water at 60° F

This equals 0 448 × latent heat of evaporation of water at 60° F

$$= 0 448 \times 1057 9$$

$$= 473 9 \text{ B Th U 's}$$

Hence the total amount of heat liberated by 11 278 lbs of furnace gas sinking from 1,481° to 60° F is 4154 1 + 473 9 = 4,628 B Th U 's This is, of course, the same as the quantity of heat absorbed in heating the same weight of gas from 60° to 1 481° F

Now from Table II, given in Chapter XI, the mean specific heat of the furnace gases between 1,481° and 3,735° F is 0 3089

Hence we have the following more exact formula for calculating the theoretical attainable temperature by the combustion gases

Let T° F be maximum flame temperature Consider the 11 278 lbs of gas to be raised first from 60° to 1 481° F and then from 1 481° to T° F

Then, as before

$$\text{Total heat} = \left\{ \begin{array}{l} \text{Heat absorbed in heating} \\ \text{combustion gas from} \\ 60^\circ \text{ to } 1,481^\circ \text{ F} \end{array} \right\} + \left\{ \begin{array}{l} \text{Heat absorbed in heating} \\ \text{combustion gas from} \\ 1,481^\circ \text{ to } T^\circ \text{ F} \end{array} \right\}$$

$$12600 + Q_c = 4628 + 11 278 (T - 1481) \times S_T,$$

or

$$T = \frac{7972 + Q_c}{11 278 \times S_T} + 1481 \quad (2)$$

Here T = flame temperature in ° F

Q = B Th U 's brought into furnace by 10 478 lbs

S_T = mean specific heat of the combustion gas between 1,481° and T° F

Now, inserting in (2) the various values of Q_c taken from Table I and the various values for S_T corresponding to the approximate flame temperatures given in Table I for various degrees of preheating the air, we then arrive at the following table (III) of more exact flame temperatures In the latter part of the table the values used for S_T are those appertaining to the calculated flame temperature just below the one required to which the temperature difference between the last pair of flame temperatures has been added This procedure

makes the results slightly more accurate in the latter part of the table than in the former, but the errors are quite negligible for practical purposes. The actual values for S_T are taken from Table II., Chapter XI. Of course the higher values of the calculated flame temperature could not be obtained in practice on account of the thermal dissociation of the various gases.

TABLE III

*More Accurate Calculations of the Theoretical Flame Temperatures
Obtainable by the Combustion Gases in a Cement Rotary Kiln*

$$T = \frac{7972 + Q}{11.278 \times S_T} + 1481.$$

| Temperature to which Entering Air is Preheated. | | Calculated Flame Temperature. | |
|--|------|-------------------------------|------|
| ° F. | ° C. | ° F. | ° C. |
| 60 | 16 | 3769 | 2076 |
| 100 | 38 | 3794 | 2090 |
| 200 | 93 | 3855 | 2124 |
| 300 | 149 | 3916 | 2158 |
| 400 | 204 | 3977 | 2192 |
| 500 | 260 | 4038 | 2226 |
| 600 | 315 | 4100 | 2260 |
| 700 | 371 | 4160 | 2293 |
| 800 | 427 | 4222 | 2328 |
| 900 | 483 | 4282 | 2361 |
| 1000 | 548 | 4344 | 2396 |
| 1100 | 594 | 4405 | 2430 |
| 1200 | 649 | 4465 | 2463 |
| 1300 | 704 | 4525 | 2496 |
| 1400 | 760 | 4586 | 2530 |
| 1500 | 815 | 4645 | 2563 |
| 1600 | 871 | 4706 | 2597 |
| 1700 | 926 | 4805 | 2652 |
| 1800 | 982 | 4862 | 2683 |
| 1900 | 1038 | 4936 | 2725 |
| 2000 | 1093 | 4995 | 2757 |
| 2100 | 1149 | 5071 | 2800 |
| 2200 | 1204 | 5147 | 2842 |
| 2300 | 1260 | 5224 | 2885 |
| 2400 | 1315 | 5301 | 2927 |
| 2500 | 1371 | 5370 | 2966 |

CHAPTER XIII

FLAME TEMPERATURES OBTAINED IN PRACTICE IN THE CEMENT ROTARY KILN

THE high theoretical flame temperatures calculated in Chapter XII. are never obtained in practice.

The following table shows the actual flame temperatures prevailing in kilns as measured by various observers :—

Observed Flame Temperatures of Rotary Kilns

| Works. | Observer. | Method of Observing. | Flame Temperature. |
|----------------------|------------------|----------------------|---------------------|
| No. 9 . . | B.P.C.R.A. * | Optical Pyrometer | 2759° F. (1513° C.) |
| " 10 . . | " | " | 2692° F. (1478° C.) |
| " 11 . . | " | " | 2536° F. (1391° C.) |
| " 26 . . | " | " | 2835° F. (1557° C.) |
| Rudersdorf (1923) . | Prof. R. Nacken† | Pt-Rh-Thermo-couple | 2606° F. (1430° C.) |
| Lehigh Valley (1905) | E. C. Soper‡ | " | 2587° F. (1420° C.) |

Of these results, probably those of Professor Nacken are the most accurate, as they were determined with the greatest possible precision, and will be taken here as the *mean flame temperature* prevailing in an ordinary rotary kiln. The optical pyrometer registers the temperature of only the hottest portion of the flame, and so tends to give high values. The theoretical flame temperature attainable with air preheated to, say, 400° F. is 3,977° F. (2,172° C.).

Reasons why the Practical Flame Temperatures Fall Short of the Theoretical Flame Temperatures

The reason why the theoretical flame temperature is not attained in a rotary kiln will be obvious from the following considerations :—

Suppose the rotary kiln is producing, say, W tons of clinker *per hour* with a coal consumption of C tons *per* 100 tons of clinker.

Suppose that a rotary kiln is consuming C tons of coal *per* 100 tons of clinker and is producing W tons of clinker *per hour*.

Then, in 1 hour $\frac{W \cdot C}{100}$ tons of coal are burnt, and in 1 sec. $\frac{W \cdot C \times 2240}{100 \times 3600}$ lbs. of coal are burnt, or 1 lb. of coal is burnt in $\frac{3600 \times 100 \times 1}{2240 \times W \cdot C}$ secs. = say, *t* secs.

* B.P.C.R.A. stands for British Portland Cement Research Association.

† Prof. R. Nacken, *Protokoll der Verhandlungen des Vereins Deutscher Portland-Cement-Fabrikanten* (E.V.). Charlottenburg, 1921, p. 184.

‡ E. C. Soper, "Report on a Test of a Portland Cement Plant," *Engineering News*, 1905, liv., Nr. 25, p. 664.

Hence, in t secs 11 278 lbs of combustion gas are produced by the burning of this coal. t is a time peculiar to each kiln working under constant circumstances, being the time in seconds required to burn 1 lb of coal. The larger the kiln, the smaller the t .

Now let us follow the history of this 11 278 lbs of gas in the clinkering zone. Suppose that the 1 lb of coal is supplied with 10 478 lbs of air preheated to, say, 400°F and placed in an insulated enclosure and allowed to burn, the time of combustion being t secs, as above described. Then under the circumstances the 11 278 lbs of gas will rise $3,977^{\circ}\text{F}$ —the theoretical flame temperature appertaining to air preheated to 400°F , as shown in Table III, Chapter XII. But in the actual kiln this 11 278 lbs of gas is losing heat for the whole t secs (by radiation, conduction, and convection), and the total amount of heat lost in this time is such as to reduce its temperature to, say, $2,600^{\circ}\text{F}$ (instead of the theoretical $3,977^{\circ}\text{F}$), as shown by Nacken.

Table II, Chapter XI, allows us to calculate this loss of heat. The instantaneous specific heat of the gases of combustion at $3,977^{\circ}\text{F}$ is $0.3426 + 0.77 \times 0.0031 = 0.3450$. The instantaneous specific heat of the gases of combustion at $2,600^{\circ}\text{F}$ is 0.3071 . So that the mean specific heat of the gases of combustion between $3,977^{\circ}$ and $2,600^{\circ}\text{F}$ is

$$\frac{0.3071 + 0.3450}{2} = 0.3260$$

Hence the amount of heat lost by radiation, conduction, and convection by 11 278 lbs of gas in t secs is

$$\begin{aligned} 11\,278 \times 0.3260 \times (3977 - 2600) \\ = 11\,278 \times 0.3260 \times 1377 \\ = 4\,836\text{ B Th U's} \end{aligned}$$

So that the percentage loss of heat on the total heat of combustion is

$$\frac{4836 \times 100}{12600 + 828} = 36.1 \text{ per cent}$$

Where is this heat going to? It is a constant loss going on steadily in a kiln. It is going to (1) maintain the walls of the kiln at their constant temperature in spite of radiation loss, i.e., compensate the radiated losses from external walls (external radiation), (2) much is radiated from the hot flame down into the cooler parts of the kiln, (3) part is communicated by conduction and convection to the material in the clinkering zone.

(1) is utterly wasteful, (2) and (3) can be usefully employed under limitations.

These questions will be dealt with later. They are of vital importance in the design of the kiln.

Since the *clinker* at the clinkering temperature tends to *evolve* heat rather than absorb it, the *practical* flame temperature is not much lowered at its initial point owing to the formation of clinker by absorption of heat, and so it comes about (as we shall see) that we can in practice calculate with a fair degree of accuracy the amount of clinker produced by a flame sinking through different temperatures.

CHAPTER XIV

WEIGHT OF CLINKER PRODUCIBLE USING COMBUSTION GASES OF DIFFERENT FLAME TEMPERATURES AND SUPPLYING 10.478 LBS. OF AIR PER 1 LB. OF STANDARD COAL BURNED

§ 1. Hot Gas is the Main Heating Agent in the Rotary Kiln.—The coal dust when fired into the kiln oxidises very rapidly into a mass of hot gas, which (as some calculations made below will show) must be considered as the main heating agent in the rotary kiln.

One lb. of standard coal yields 11.278 lbs. of hot combustion gas, which, in the ordinary cement kiln, starts at an initial temperature of about 2,600° F. (1,427° C.) and travels down the kiln and imparts its heat to the raw material in the kiln.

1,481° F.

| Above 1,481° F. | : Below 1,481° F. |
|--|-------------------|
| A. Clinkering and Decarbonating Zone. | B. |

FIG. 1.

§ 2. The kiln may be divided into two parts—a portion A, where the temperature inside the kiln is *above* 1,481° F. (805° C.), and a portion B, where the zone is below 1,481° F. (805° C.) (see Fig. 1).

1,481° F. (805° C.) is the temperature whereat calcium carbonate begins to decompose in the furnace, and no clinker can be formed until the chalk does decompose.

So that any heat escaping past the *line* dividing the A portion of the kiln from the B portion is completely lost so far as actual cement formation is concerned—the heat thus escaping being merely employed in preheating and drying the raw material preparatory for its conversion into clinker.

§ 3. The quantity of clinker formed, therefore, is measured by the amount of heat *absorbed* by the raw material between 1,481° F. (805° C.) and the clinkering temperature (2,498° F.). This quantity of heat *Q* must, in the ideal case, be equal to the amount of heat given out by the hot gas between the time of

entering the kiln at *mn* and leaving it at *op* (since no radiation and conduction loss occurs) (Fig. 2)

If the combustion gas from 1 lb of coal starts in the section A at its maximum temperature of $T^{\circ}\text{F}$, and leaves the section A at the temperature $1,481^{\circ}\text{F}$, the amount of heat it has given up by the 11 278 lbs of gas passing down the section A is

$$Q = 11\,278\ S\ (T - 1481)\ \text{B Th U 's},$$

where S = the specific heat of the gas

To form 1 lb of clinker requires the absorption of 918.6 B Th U 's by the raw material between $1,481^{\circ}\text{F}$ and its clinkering temperature (see Chapter IV) So that the weight in lbs of clinker formed by the 11 278 lbs of hot gas (= 1 lb of standard coal) in the section A is

$$W = \frac{Q}{918.6} = \frac{11\,278\ S\ (T - 1481)}{918.6} \quad (1)$$

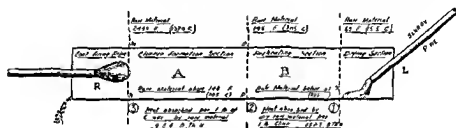


FIG. 2

For some purposes the following formula is somewhat more convenient for use in calculating W , although, in fact, it is the same formula as (1)

If W is the number of lbs of clinker producible by 1 lb of standard coal—

$$W = \frac{\text{B Th U 's contained in combustion gases from 1 lb coal above } 1,481^{\circ}\text{F}}{\text{B Th U 's necessary to form 1 lb of clinker above } 1,481^{\circ}\text{F.}}$$

$$= \frac{\left\{ 12,600 + \text{B Th U 's in } 10\,468\ \text{lbs of preheated air} \right\} - \left\{ \text{B Th U 's in combustion gas between } 1,481^{\circ}\text{ and } 60^{\circ}\text{F} \right\}}{918.6}$$

§ 4 In order to connect the rate of formation of the clinker with the amount of coal burnt, we proceed as follows —

In t secs the kiln consumes 1 lb of standard coal, and therefore produces 11 278 lbs of combustion gas. Hence in t secs 11 278 lbs of combustion gas enter the section A at temperature $T^{\circ}\text{F}$, and 11 278 lbs simultaneously leave it at a temperature $1,481^{\circ}\text{F}$, the heat, therefore, lost by the hot gas in t secs in the section A is equal to the heat lost by 11 278 lbs of gas sinking from T° to $1,481^{\circ}\text{F}$. So that in t secs the amount of heat absorbed by the raw material in A is

$$Q = 11\,278\ S\ (T - 1481)\ \text{B Th U 's},$$

and therefore the amount of clinker produced in t secs is

$$W = \frac{11\,278\ S\ (T - 1481)}{918.6}\ \text{B Th U 's} \quad (1)$$

So that the amount of clinker produced by the kiln in 1 sec is

$$\frac{W}{t} = \frac{11\,278\ S\ (T - 1481)}{918.6\ t}\ \text{lbs} \quad (2)$$

and the tonnage producible per hour by the kiln is

$$\frac{11.27S \cdot S \cdot (T - 1481) \times 3600}{918.6 \times 2400} \text{ tons} \quad (3)$$

§ 5. The next step is to calculate the amount of heat which can be liberated by the 11.27S lbs. of the combustion gas (=1 lb. of standard coal) in sinking from various flame temperatures down to 1,481° F.

On dividing this amount of heat by 918.6 (representing the number of B.Th.U.'s which must be absorbed by the raw material between 1,481° F. and the clinkering temperature in order to produce 1 lb. of clinker) we get the number of lbs. of clinker producible by 1 lb. of standard coal under these circumstances.

These calculations are carried out in the following table (I.).

The various theoretical flame temperatures tabulated in Table III., Chapter XVII., are placed in column 1.

TABLE I

| (1) A Theoretical Maximum Flame Temperature (as Calculated in Table III., Chapter XII.). | | (2) B.Th.U.'s Liberated by 11.27S Lbs. of Combustion Gas Falling from T° to 1,481° F. | (3) Lbs. of Clinker Theoretically Pro- ducible per 1 Lb. of Standard Coal of 12,600 B.Th.U.'s Burnt in Upper Part of Kiln. | (4) Tons of Standard Coal Consumed per 100 Tons Clinker Produced. | (5) Temperature of Preheated Air for Coal Combustion. | |
|--|------|---|---|---|---|------|
| T. | | $Q = 11.27S \times S \times (T - 1481).$ | $\frac{Q}{918.6} = W.$ | $\frac{918.6 \times 100}{Q}$ | t°. | |
| ° F. | ° C. | | | | ° F. | ° C. |
| 5370 | 2966 | 14,452 | 15.732 | 6.36 | 2300 | 1371 |
| 5301 | 2927 | 14,160 | 15.414 | 6.49 | 2400 | 1315 |
| 5224 | 2883 | 13,873 | 15.102 | 6.62 | 2300 | 1260 |
| 5147 | 2842 | 13,589 | 14.793 | 6.76 | 2200 | 1204 |
| 5071 | 2800 | 13,306 | 14.485 | 6.90 | 2100 | 1149 |
| 4995 | 2757 | 13,023 | 14.177 | 7.03 | 2000 | 1093 |
| 4936 | 2725 | 12,744 | 13.873 | 7.21 | 1900 | 1038 |
| 4862 | 2683 | 12,470 | 13.575 | 7.37 | 1800 | 982 |
| 4805 | 2652 | 12,191 | 13.271 | 7.53 | 1700 | 926 |
| 4706 | 2597 | 11,915 | 12.972 | 7.71 | 1600 | 871 |
| 4645 | 2563 | 11,643 | 12.675 | 7.89 | 1500 | 815 |
| 4586 | 2530 | 11,378 | 12.386 | 8.07 | 1400 | 760 |
| 4525 | 2496 | 11,110 | 12.094 | 8.27 | 1300 | 704 |
| 4463 | 2463 | 10,845 | 11.806 | 8.47 | 1200 | 649 |
| 4405 | 2430 | 10,584 | 11.522 | 8.68 | 1100 | 594 |
| 4344 | 2396 | 10,324 | 11.239 | 8.90 | 1000 | 548 |
| 4282 | 2361 | 10,062 | 10.954 | 9.13 | 900 | 483 |
| 4222 | 2328 | 9,806 | 10.675 | 9.37 | 800 | 427 |
| 4160 | 2293 | 9,551 | 10.397 | 9.62 | 700 | 371 |
| 4100 | 2260 | 9,301 | 10.125 | 9.88 | 600 | 315 |
| 4038 | 2226 | 9,048 | 9.849 ⁸ | 10.15 | 500 | 260 |
| 3977 | 2192 | 8,800 | 9.580 | 10.44 | 400 | 204 |
| 3916 | 2158 | 8,553 | 9.311 | 10.74 | 300 | 149 |

TABLE I --(Continued)

| (1) A Theoretical Maximum Flame Temperature (as Calculated in Table III Chapter XII) | | (2) B Th U 's Liberated by 11 2/3 Lbs of Combustion Gas Falling from T° to 1,451° F | (3) Lbs of Chinker Theoretically Pro- ducible per 1 Lb of Standard Coal of 12 600 B Th U 's Burnt in Upper Part of Kilm | (4) Tons of Standard Coal Consumed per 100 Tons Chinker Produced | (5) Temperature of Preheated Air for Coal Combustion | |
|--|------|---|--|--|--|----|
| T | | $Q = 11.2, 8 \times 5$ $\times (T - 1,451)$ | $\frac{Q}{915.6} = W$ | $\frac{915.6 \times 100}{Q}$ | t° | |
| °F | °C | | | | °F | °C |
| 3855 | 2124 | 8309 | 9.045 | 11.06 | 200 | 93 |
| 3794 | 2090 | 8070 | 8.785 | 11.38 | 100 | 38 |
| 3769 | 2076 | 7972 | 8.680 | 11.52 | 60 | 16 |
| 3700 | 2038 | 7718 | 8.408 | 11.89 | | |
| 3600 | 1982 | 7334 | 7.984 | 12.52 | | |
| 3500 | 1926 | 6954 | 7.570 | 13.21 | | |
| 3400 | 1872 | 6582 | 7.165 | 13.96 | | |
| 3300 | 1816 | 6210 | 6.761 | 14.79 | | |
| 3200 | 1760 | 5841 | 6.359 | 15.73 | | |
| 3100 | 1704 | 5478 | 5.963 | 16.77 | | |
| 3000 | 1649 | 5115 | 5.568 | 17.96 | | |
| 2900 | 1594 | 4759 | 5.181 | 19.30 | | |
| 2800 | 1538 | 4405 | 4.795 | 20.85 | | |
| 2700 | 1482 | 4053 | 4.413 | 22.66 | | |
| 2600 | 1427 | 3704 | 4.032 | 24.80 | | |
| 2590 | 1421 | 3668 | 3.993 | 25.04 | | |
| 2580 | 1415 | 3634 | 3.956 | 25.28 | | |
| 2570 | 1410 | 3600 | 3.919 | 25.52 | | |
| 2560 | 1404 | 3566 | 3.882 | 25.76 | | |
| 2550 | 1399 | 3530 | 3.843 | 26.02 | | |
| 2540 | 1393 | 3496 | 3.806 | 26.27 | | |
| 2530 | 1388 | 3462 | 3.769 | 26.53 | | |
| 2520 | 1382 | 3427 | 3.731 | 26.80 | | |
| 2510 | 1377 | 3392 | 3.693 | 27.08 | | |
| 2500 | 1371 | 3358 | 3.656 | 27.35 | | |
| 2490 | 1365 | 3324 | 3.619 | 27.63 | | |
| 2480 | 1360 | 3290 | 3.582 | 27.92 | | |
| 2470 | 1354 | 3256 | 3.545 | 28.21 | | |
| 2460 | 1349 | 3221 | 3.507 | 28.51 | | |
| 2450 | 1343 | 3221 | 3.460 | 28.83 | | |
| 2440 | 1337 | 3183 | 3.432 | 29.14 | | |
| 2430 | 1332 | 3119 | 3.395 | 29.45 | | |
| 2420 | 1327 | 3085 | 3.359 | 29.77 | | |
| 2410 | 1321 | 3050 | 3.320 | 30.12 | | |
| 2400 | 1315 | 3016 | 3.283 | 30.46 | | |
| 2300 | 1260 | 2677 | 2.915 | 34.30 | | |
| 2200 | 1204 | 2341 | 2.549 | 39.25 | | |
| 2100 | 1149 | 2007 | 2.185 | 45.77 | | |
| 2000 | 1093 | 1676 | 1.821 | 54.81 | | |
| 1900 | 1038 | 1347 | 1.466 | 70.23 | | |
| 1800 | 982 | 1011 | 1.111 | 90.00 | | |
| 1700 | 926 | 698 | 0.760 | 131.58 | | |
| 1600 | 871 | 378 | 0.411 | 243.31 | | |
| 1500 | 815 | 60 | 0.065 | 1538.46 | | |
| 1481 | 805 | 0 | 0.00 | Infinite | | |

The mean specific heat S of the furnace gases between T° and $1,481^\circ$ is next taken from *Table II.*, Chapter XI. Then the required quantity of heat in column 2 is calculated by the formula

$$Q = 11.278 \times S \times (T - 1481^\circ).$$

Column 3 then gives the weight of clinker theoretically producible per 1 lb. of coal burnt in the upper part of the kiln, which, for convenience for the manufacturer, is reduced to tons of standard coal consumed per 100 tons of cement in the next column (4).

For convenience' sake I have added in column 5 the temperature to which the entering air must be preheated in order to obtain the flame temperatures indicated in column 1.

A study of Table I. leads to the following important conclusions:—

(1) An ideal kiln, subject to no losses of any kind, could produce 100 tons of cement clinker by the combustion of 6.36 tons of standard coal.

The ordinary wet process rotary kiln is supposed to be doing well when it produces 100 tons of clinker for 28 tons of standard coal, so that the efficiency of the ordinary kiln is only

$$\frac{6.36 \times 100}{28} = 22.7 \text{ per cent.}$$

There is, therefore, much room for improvement in the rotary kiln.

(2) In an actual rotary kiln the *practically observed flame temperatures correspond with the theoretical output of clinker appertaining to that temperature.* Thus the observed flame temperature of $2,600^\circ \text{ F.}$ corresponds with a clinker output of 100 tons per 24.80 tons coal fired.

This has actually been achieved in kilns belonging to certain works, and means that the gases, starting at $2,600^\circ \text{ F.}$, actually give up all *their effective heat* to the raw material (*i.e.*, the gases escape from the decarbonating zone at a temperature degree of $1,481^\circ \text{ F.}$, so that they have given up all their effective heat to the raw material). In most kilns, however, the output is 28 to 33 tons of coal per 100 tons clinker, the *reason being* that the combustion gases *travel so quickly down the kiln* that they escape from the *decarbonating zone* not at $1,481^\circ \text{ F.}$, but at, say, $1,700^\circ \text{ F.}$, so that the effective heat has not been communicated to the raw material. This is dealt with in the next chapter.

(3) The fact that the *practically observed flame temperatures agree with clinker output in practice leads to the most important practical conclusion that the heating in the kiln is indeed carried out by the hot gas and is measured by the heat loss from the gas as it passes down the kiln.*

The importance of this conclusion is this: It follows that if you wish to increase the yield of clinker much beyond 100 tons per 24 tons of standard coal, *it is absolutely vital to increase the flame temperature in the clinkering zone.*

From equation (1), if W is the weight of clinker formed per 1 lb. of coal burnt and T is the flame temperature—

$$W = \frac{11.278 \cdot S \cdot (T - 1481)}{918.6},$$

or

$$W = \text{constant} \times (T - 1481).$$

Hence the higher is T , the *larger* is W .

How the flame temperature can be increased without ruining the lining will be explained in a later chapter.* It forms a vital element in the design of a kiln.

* In Vol. II. of this Report, not yet written because the British Portland Cement Research Association was dissolved before the work could be completed.

(4) *The higher the flame temperature of the gaseous mixture, the greater the clinker output per 1 lb of coal burnt*

This is a conclusion of great practical importance. Any influence which diminishes the flame temperature of the gases necessarily diminishes the clinker output per 1 lb of coal burnt.

The subject will be treated again in a later chapter, and the influence of excess of air on clinker output will be examined and shown to be deleterious.

CHAPTER XV

FOR FUEL ECONOMY IT IS ESSENTIAL TO
UTILISE HIGH-GRADE HEAT FOR HIGH-
GRADE THERMAL WORK AND LOW-GRADE
HEAT FOR LOW-GRADE THERMAL WORK

The Entropy of Portland Cement Formation

§ 1. **Distinction between High-Grade and Low-Grade Heat.**—It is essential to grasp the difference between high-grade and low-grade heat. Neglect of the fact that 1 B.Th.U. when liberated at a high temperature can do very different amounts of thermal work than when liberated at low temperatures has caused the heat balance of the rotary kiln to be much misunderstood, and very misleading practical consequences have as a result been deduced therefrom, which have led to serious errors in kiln design.

§ 2. In order to illustrate the difference between high-grade and low-grade heat let us consider 11.278 lbs. of combustion gas produced by burning 1 lb. of standard coal with 10.478 lbs. of air (see Chapter X.), which occurs in the ordinary cement rotary kiln. If, as in the ordinary kiln, the flame temperature of the resulting gas is $2,600^{\circ}\text{F}$., then 11.278 lbs. of combustion gas falling in temperature from $2,600^{\circ}$ to 60°F . can liberate 8,332 B.Th.U.'s.

Now mix this hot gas with 10.727 lbs. of air at 60°F . Then the resulting mixture will acquire a temperature of $1,481^{\circ}\text{F}$. If now the mixture of 22.005 lbs. of gas be allowed to fall in temperature from $1,481^{\circ}$ to 60°F ., then precisely the same amount of heat as before will be liberated, viz., 8,332 B.Th.U.'s. So that 11.278 lbs. of furnace gas falling in temperature from $2,600^{\circ}$ to 60°F . will evolve exactly the same number of B.Th.U.'s as 22.005 lbs. of mixed gas falling from $1,481^{\circ}$ to 60°F ., viz., 8,332 B.Th.U.'s.

§ 3. But these two lots of 8,332 B.Th.U.'s, although equal in number, are not identical in properties. There is this vital difference between them—the 8,332 B.Th.U.'s from the furnace gas can produce 3.993 lbs. of clinker (as shown in Chapter XIV., Table I.), whereas the 8,332 B.Th.U.'s from the mixed gas are

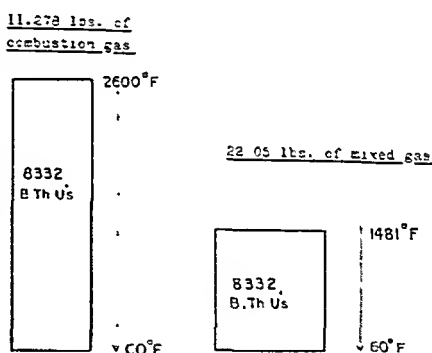


FIG. 1.—Illustrating the fact that the same number of B.Th.U.'s can be obtained from 11.278 lbs. of Furnace Gas sinking from $2,600^{\circ}\text{F}$. as from 22.05 lbs. of Mixed Gas sinking from $1,481^{\circ}$ to 60°F .

unable to produce a single ounce of clinker. The first lot of B Th U's consist of what we term high grade heat, while the other lot consist of low grade heat. Hence high grade heat does not act in the same way as low grade heat.

A certain number of B Th U's available at a high temperature will act entirely different from, and possess an entirely different value from, the same number of B Th U's which are only available at a low temperature. There is as much difference between one unit of high grade heat and one unit of low grade heat as there is between 1 lb of diamonds and 1 lb of coal. In the first case the same amount of heat is available, but its quality is different. In the second case the same amount of carbon may be available, but its quality is different.

§ 4 Now let us go back and consider in greater detail the nature of the B Th U's evolved from the 11 278 lbs of furnace gas.

As explained in Chapter XIV, the high grade heat available for clinker forma-

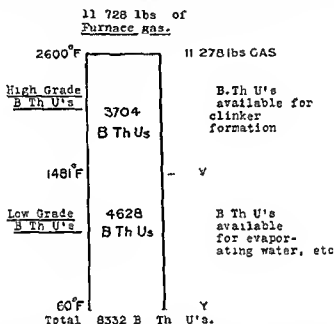


FIG 2—Illustrating the number of High grade and Low grade B Th U's contained in 11 278 lbs of Furnace Gas heated to 2600° F (the ordinary Flame Temperature prevailing in the Rotary Kiln)

tion consists of those B Th U's which are evolved by the gas above the temperature level of 1,481° F (the temperature of decomposition of the CaCO_3 in the kiln).

Hence in this case since 3,704 B Th U's are evolved by 11 278 lbs of furnace gas sinking from 2600° to 1,481° F out of the total of 8,332 B Th U's contained in the gas, only 3,704 are available for clinker formation and the remaining 4,628 B Th U's (which are evolved between 1,481° and 60° F) can only be used in the dehydrating and preheating zone for evaporating water and the like.

§ 5 Hence we have this vitally important fact. Although the whole 8,332 B Th U's contained in the 11 268 lbs of combustion gas between 2600° and 60° F can be utilised for evaporating water and preheating slurry, if we so desire, it is impossible to employ them all in making clinker. Only the 3,704 B Th U's can be employed for making clinker and therefore, so far as the cement manufacturer is concerned, these 3,704 B Th U's are especially valuable, whereas the remaining 4,628 B Th U's possess little value.

§ 6. If, therefore, any of these valuable 3,704 B.Th.U.'s are employed in doing the low-grade thermal work of evaporating water or preheating slurry, we are using valuable high-grade heat in doing inferior thermal work, which can be done equally well, if not better, by the more abundant and cheaper low-grade heat.

It is as if we are employing a valuable race-horse for drawing a plough instead of employing the much cheaper cart-horses for the job, or as if we are throwing diamonds into a furnace to develop heat instead of employing ordinary coal.

§ 7. This, then, is what happens when the furnace gases are allowed to escape from the decarbonating zone at any temperature above $1,481^{\circ}\text{F}$. ($1,371^{\circ}\text{C}$). They are carrying with them unutilised a certain proportion of high-grade heat, which is then utilised in evaporating water in the lower part of the kiln. Let us illustrate this by a concrete case.

Suppose that the furnace gases in the kiln escape from the decarbonating zone into the dehydrating zone at a temperature of, let us say, $1,800^{\circ}\text{F}$. Then the

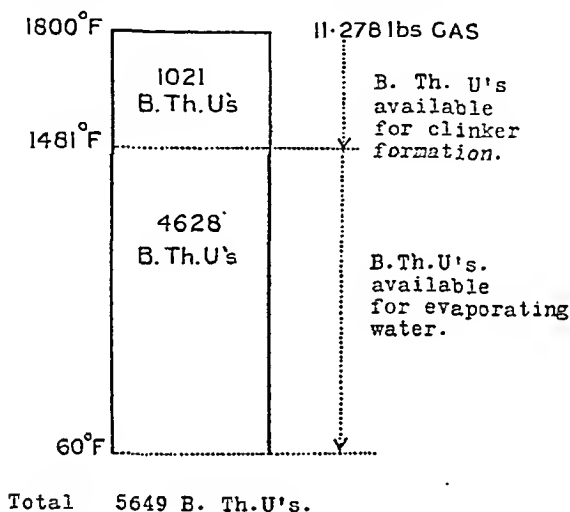


FIG. 3.

total heat contained in 11.268 lbs. of the combustion gas between $1,800^{\circ}$ and 60°F . is 5,649 B.Th.U.'s, *all of which* is available for evaporating the water and drying the slurry. Yet of these 5,649 B.Th.U.'s only 1,021 are available for forming clinker, and therefore, if these 1,021 B.Th.U.'s are allowed to escape into the evaporating zone and be employed in evaporating water instead of liberating clinker, it is obvious that a very serious loss of high-grade heat is going on (see Fig. 3).

This loss is all the more serious because if the above-mentioned 1,021 B.Th.U.'s were employed in forming clinker, they would liberate at $1,481^{\circ}\text{F}$. no less than 0.51 lb. of CO_2 from the calcium carbonate, so that the combined mass of gas passing down the kiln would still have available for evaporation of water and preheating the slurry all the 1,021 B.Th.U.'s after this latter had performed their due work of forming clinker.

All that would have happened to the 1,021 B.Th.U.'s after performing the thermal work of producing clinker would be that it had become transformed from the category 1,021 B.Th.U.'s of high-grade heat (*i.e.*, heat available above $1,481^{\circ}\text{F}$.) into 1,021 B.Th.U.'s of lower-grade heat.

§ 8 It is, therefore, an absolute and irretrievable wastage to allow any high grade heat at all to escape into the lower part of the kiln without forming from it all the clinker we can

It is as if we desired to obtain work from a high pressure steam boiler containing steam at 250 lbs pressure, and yet before allowing the steam to pass into the cylinders of the engine we deliberately allowed the pressure of the steam to fall to, say, 50 lbs by passing it through an expansion valve. All engineers would condemn this practice, yet it is precisely similar to what cement manufacturers are doing when they allow high grade heat from the upper part of the kiln to pass uselessly down into the lower part without abstracting from it all its available capacity for producing clinker before it reaches the evaporating zone

§ 9 It is this neglect to trap the high grade heat in the right part of the kiln which makes the ordinary cement kiln so wasteful and inefficient as a thermal engine

The whole attention of cement manufacturers should be concentrated on keeping the high grade heat in its proper sphere, viz., in the clinkering and decarbonating zone, and not allowing any to escape and be wasted in the dehydrating and pre-heating zone

This, as will be proved in Chapter XVII, is only possible if the flame temperature of the gases is maintained as high as possible in the clinkering zone and as much heat as possible abstracted from the hot gases in the decarbonating zone

Development of the Notion of High grade Heat

On account of the vital importance of this aspect of cement manufacture we will treat the subject in a different way, so as to bring into ordinary engineering practice this conception of *high grade heat* and *low grade heat*

When we are dealing with furnaces in which substances are made to undergo chemical or physical changes, it is not only the quantity of heat which counts, but even more so the thermal pressure (or temperature, as physicists call it) at which the heat is delivered to the substance. For example, 100 British units of heat available at a thermal pressure of 212°F are much less valuable than 100 units of heat available at, say, $2,600^{\circ}\text{F}$. In the one case 100 units of heat will produce no cement clinker at all, and in the other case the 100 units can be converted into an equivalent amount of clinker

The ordinary thermal balance of the rotary kiln is practically useless from a technical point of view, because the quantities of heat are all expressed as so many B Th U's without distinguishing whether these B Th U's are available at a low thermal pressure or temperature.

We have just seen, for example, in Chapter XIV, that when the flame temperature in the kiln is at its maximum of $5\,370^{\circ}\text{F}$ ($2,966^{\circ}\text{C}$), we can produce 100 tons of cement clinker by the expenditure of only 6.36 tons of standard coal (12 600 B Th U's per lb. One ton = 2,240 lbs = 1.12 American tons)

Whereas with a flame temperature of $2,600^{\circ}\text{F}$ we could produce 100 tons of clinker by the combustion of 24.8 tons of standard coal, and with a flame of only $1,481^{\circ}\text{F}$ (805°C) it is necessary to burn an infinite amount of coal in order to produce an ounce of clinker

The Case Analogous to the Generation of Steam in a Boiler—Millions of B Th U's passed through the heating flues of a boiler below the boiling point of water (100°C or 212°F at atmospheric pressure) will not generate steam. The elevation of the temperature of the heating gases above 212°F is what decides the value of the heating medium so far as steam raising is concerned

In the same way, in cement production the value of the gaseous heating medium for producing cement clinker is measured by the number of B Th U's available

above 805° C. (1,481° F.)—the temperature of decomposition of the calcium carbonate, CaCO_3 , the temperature whereat the great heat absorption occurs in cement manufacture.

The Entropy of Cement Formation

Mathematical physicists express the fact that not only is the quantity of heat important, but also the thermal pressure, or temperature, at which it is delivered by introducing the conception of "*entropy*." The change of entropy of a substance is measured by the quantity of heat which passes into the substance divided by the absolute temperature at which it passes in.

In mathematical language: If ϕ be the change of entropy, Q the quantity of heat, and T the absolute temperature, then

$$\phi = \int \frac{dQ}{T}.$$

The units of entropy are termed "ranks," after the celebrated engineer Rankine, who largely developed the conception of entropy as applied to steam engine and boiler design.

Indeed, the conception of entropy is of prime importance to the steam engineer. By its means problems in steam engineering are easily and rapidly solved, which would be difficult, if not impossible, to solve by any other method.

The fundamental advances which have been made in the designing of furnace boilers and steam engines within modern times largely sprang from the conception of entropy and available energy. The magnificent entropy tables, published by Prof. Callendar and others within recent years, have made accessible to all steam engineers the necessary data. Of no less importance for correct cement-kiln design is a knowledge of the magnitude of an entropy, changes which occur at different stages of cement-clinker formation.

Calculations made in Chapter XIV. on the thermodynamics of cement formation showed that by suitable thermodynamic design an ideal kiln could produce 100 tons of cement clinker, by the combustion of 6.36 tons of standard coal (12,600 B.Th.U.'s per lb.), and that by modifying the design of kilns to conform to the changes of entropy in the raw material in different parts of the kiln, kilns very much more efficient than any now existing could be constructed. Consequently practical kiln designers, familiar with thermodynamic theory, will welcome the table on following page in which I have calculated the change of entropy of the solid materials in making 1 lb. of Portland-cement clinker, reckoned from 32° F. as zero. As far as I am aware, this is the first time that such tables have been calculated for the cement industry. No doubt more accurate tables will in future be made possible when the thermal data relating to cement formation has been more accurately determined.*

Application of the Preceding Principles to the Thermal History of Portland Cement Formation

Returning now to practical details, in order to make Portland cement we must heat a mixture of clay and calcium carbonate (chalk, limestone, etc.).

But as we have seen, the heat must be supplied in a definite way. We must not only supply a certain *quantity of heat*, but we must also supply it at the right thermal pressure (or temperature, as physicists say).

In order to see this, let us now follow the thermal history of a batch of raw material through the kiln.

As water plays no part in the formation of cement, and is a mere mechanical

* The table was first published in *Cement and Cement Manufacture*, and is reproduced here by kind permission of the Editor.

TABLE I

Change of Entropy of the Solid Materials which occur in Making 11 Lbs of Cement Clinker, Reckoned from 32° F as Zero

| Temperature | | Entropy Ranks | Remarks | Total Change of Entropy during Different Stages | Total Change of Entropy during Different Zones |
|-------------|------|--|--|---|--|
| ° F | ° F | ϕ | | | |
| 32 | 492 | 0 00 | { Evolution of water from hydrated silica } | 0 128 | { Drying zone, 0 124 } |
| 212 | 872 | 0 124 | | | |
| 212 | 672 | { 0 124 at beginning of 212° F 0 130 at end of 212° F } | | | |
| 212 | 772 | | | | |
| 312 | 772 | 0 189 | { Evolution of water from kaolin } | 0 006 | |
| 412 | 872 | 0 238 | | | |
| 512 | 972 | 0 287 | | | |
| 612 | 1072 | 0 322 | | | |
| 712 | 1172 | 0 358 | | | |
| 812 | 1272 | 0 391 | | | |
| 912 | 1372 | 0 427 | | | |
| 1012 | 1472 | 0 450 | | | |
| 1112 | 1572 | { 0 477 at beginning of 1112° F 0 486 at end of 1112° F } | | | |
| 1212 | 1672 | 0 510 | | | |
| 1312 | 1772 | 0 534 | { Evolution of Co ₂ from calcium carbonate } | 0 009 | |
| 1412 | 1872 | 0 556 | | | |
| 1481 | 1872 | 0 591 | | | |
| 1481 | 1941 | { 0 571 at beginning of 1481° F 0 989 at end of 1481° F } | | | |
| 1481 | 1941 | 0 989 | | | |
| 1581 | 2041 | 1 0033 | | | |
| 1681 | 2141 | 1 0168 | | | |
| 1781 | 2241 | 1 0297 | | | |
| 1881 | 2341 | 1 0420 | | | |
| 1981 | 2441 | 1 0539 | | | |
| 2081 | 2541 | 1 0651 | { Evolution of heat due to exothermic reaction of clinker for matron } | 0 418 | { Decarbonating zone, 0 418 } |
| 2181 | 2641 | 1 0761 | | | |
| 2281 | 2741 | 1 0864 | | | |
| 2381 | 2841 | 1 0967 | | | |
| 2481 | 2941 | 1 1063 | | | |
| 2498 | 2958 | 1 1080 | | | |
| 2498 | 2958 | { 1 1080 at beginning of 2498° F 1 0472 at end of 2498° F } | | | |
| 2498 | 2958 | | | | |
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admixture, it simplifies the treatment of the subject to consider that the mixture is dry at a temperature of 60° F. (15.6° C.).

Assume that this batch of raw material consists of:

- 1.1905 lb. of calcium carbonate (CaCO_3).
- 0.1846 lb. of kaolin ($\text{Al}_2\text{Si}_2\text{O}_7 \cdot 2\text{H}_2\text{O}$).
- 0.1540 lb. of hydrated silica (93 per cent. SiO_2 , 7 per cent. H_2O).
- 0.0312 lb. of ferric oxide (Fe_2O_3).

Total: 1.5603 lb.

This mixture, when ignited to about 2498° F. (1370° C.) will form 1 lb. of cement clinker of the ordinary type manufactured in Great Britain.

As the temperature (or thermal pressure) of this 1 lb. mass of raw material increases—*i.e.*, as it becomes hotter and hotter—different quantities of heat are absorbed, or, so to speak, pumped into the material at different temperature levels (or better, at different thermal pressures).

The following table shows this, which I have calculated from data compiled by the British Portland Cement Research Association:—

TABLE II

Quantity of Heat Absorbed by the Solid Raw Materials in Making 1 Lb. of Portland Cement Clinker between 32° F. and the Clinkering Temperature $2,498^{\circ}$ F.

| $^{\circ}$ F. | Quantity of heat in B.Th.U.'s. | |
|---------------|---|--|
| 32 | 0.00 | |
| 212 | $\left\{ \begin{array}{l} 70.29 \text{ beginning of } 212^{\circ} \text{ F.} \\ 77.09 \text{ end of } 212^{\circ} \text{ F.} \end{array} \right.$ | $\left\{ \begin{array}{l} \text{Due to splitting off} \\ \text{of water from hy-} \\ \text{drated silica} \end{array} \right\} 70.29 \text{ B.Th.U.'s in drying zone.}$ |
| 312 | 117.60 | } 534.20 B.Th.U.'s in preheating zone. |
| 412 | 158.04 | |
| 512 | 198.59 | |
| 612 | 239.02 | |
| 712 | 279.57 | |
| 812 | 320.01 | |
| 912 | 360.54 | |
| 1012 | 400.89 | } 811.82 B.Th.U.'s absorbed owing to evolution of CO_2 by CaCO_2 during zone. |
| 1112 | $\left\{ \begin{array}{l} 441.42 \text{ at beginning of } 1112^{\circ} \text{ F.} \\ 455.32 \text{ at end of } 1112^{\circ} \text{ F.} \end{array} \right.$ | $\left\{ \begin{array}{l} \text{Due to splitting off} \\ \text{of water from kaolin} \end{array} \right.$ |
| 1212 | 495.68 | } 286.70 B.Th.U.'s absorbed in heating the raw materials up to 2498° F. and 179.93 B.Th.U.'s are evolved when chemical union takes place, making net absorption 106.77 B.Th.U.'s. |
| 1312 | 536.08 | |
| 1412 | 576.47 | |
| 1481 | $\left\{ \begin{array}{l} 604.49 \text{ at beginning of } 1481^{\circ} \text{ F.} \\ 1416.41 \text{ at end of } 1481^{\circ} \text{ F.} \end{array} \right.$ | |
| 1581 | 1444.61 | |
| 1681 | 1472.81 | |
| 1781 | 1501.01 | |
| 1881 | 1529.21 | } Sintering zone |
| 1981 | 1557.41 | |
| 2081 | 1585.61 | |
| 2181 | 1613.71 | |
| 2281 | 1641.91 | |
| 2381 | 1670.21 | |
| 2481 | 1698.31 | |
| 2498 | $\left\{ \begin{array}{l} 1703.11 \text{ at beginning of } 2498^{\circ} \text{ F.} \\ 1523.18 \text{ at end of } 2498^{\circ} \text{ F.} \end{array} \right.$ | $\left\{ \begin{array}{l} \text{Due to exothermic} \\ \text{reaction.} \end{array} \right.$ |

In Making Cement Clinker it is Necessary to Force Different Quantities of Heat (B Th U 's) at Different Thermal Pressures (or Temperatures) into the Raw Materials

By studying this table you will see that between 32° and 212° F, the raw material must have forced into it a quantity of heat measured by 70.29 B Th U 's. So that from a practical point of view, 70.29 B Th U 's must be pumped into the raw material at a *thermal pressure not below 212° F*

Next, between 212° and $1,481^{\circ}$ F, you will see that the material must have forced into it an additional 534 B Th U 's. So that the kiln designer must arrange matters that this number of B Th U 's must be available from the surrounding heating medium at a thermal pressure between 212° and $1,481^{\circ}$ F

Next, at $1,481^{\circ}$ F—the temperature at which the calcium carbonate begins to decompose and evolve carbon dioxide—the material must have pumped into it no less than 812 B Th U 's delivered at a thermal pressure above $1,481^{\circ}$ F

If these B Th U 's are available from the heating medium at a *lower* thermal pressure than $1,481^{\circ}$ F, they are of no practical value so far as decomposing the calcium carbonate and expelling the CO_2 therefrom is concerned. At a lower thermal pressure than $1,481^{\circ}$ F the B Th U 's simply cannot be pumped into or absorbed by the material, and they pass away through the kiln unabsorbed and without doing any useful chemical work, and escape up the chimney to the air unutilised, merely increasing the exit temperature of the kiln in so doing.

Finally, about 107 B Th U 's must be forced into the material between $1,581^{\circ}$ and $2,498^{\circ}$ F in order to cause the lime and silica to chemically unite to form the mixture of calcium silicates, known as Portland cement clinker. And if these 167 B Th U 's are not available from the heating medium at a thermal pressure of about $2,500^{\circ}$ F, then no cement clinker is formed, no matter how many millions of B Th U 's are forced through the kiln or how many tons of coal are burnt.

This fact was strikingly illustrated some years ago by costly large scale experiments, carried out by the Associated Portland Cement Manufacturers, at their Swanscombe Works, by Eldred. He obtained hot gas derived from a producer whose flame temperature was slightly lower than $2,500^{\circ}$ F and forced it through a rotary kiln fed with slurry. The material came out severely under burnt, and the exit temperature at the end of the kiln shot up, while the fuel consumption increased enormously.

These experiments cost nearly £5,000, and demonstrated the impossibility of making cement by these methods.

CHAPTER XVI

ON THE PREHEATING OF THE ENTERING AIR BY MEANS OF HOT CLINKER

§ 1. THE temperature at which clinker issues from the ordinary cement kiln is about 2,500° F.

The following figures shown in Table I. have been ascertained regarding the clinkering temperature of different kilns by different observers at different times.

No doubt different mixtures have different clinkering temperatures.

In the following calculations the temperature of the clinker is taken as 2,500° F. (1,371° C.):—

TABLE I

*Observed Temperatures of Clinker at the Clinkering Temperature
in the Rotary Kiln*

| Works. | Observer. | Method of Observing. | Temperature of Clinker. |
|----------------------|----------------|----------------------|--|
| No. 9 . . . | B.P.C.R.A.* | Optical Pyrometer | 2554° F. (1401° C.) |
| „ 10 . . . | „ | „ | 2447° F. (1342° C.) |
| „ 11 . . . | „ | „ | 2298° F. (1259° C.) |
| „ 26 . . . | „ | „ | 2444° F. (1340° C.) |
| Rudersdorf (1923) . | J. C. Gullen † | Seeger Cones | 2552° F. (1400° C.) |
| | R. Nacken ‡ | Pt-Rh-Thermo-couple | 2498° F. (1370° C.) |
| Lehigh Valley (1905) | E. C. Soper § | „ | 2496° F. (1369° C.) |
| Rudersdorf . . . | H. Kühl | ... | 2552° F. (1400° C.) |
| | O. Schott ¶ | ... | 2552° - 2642° F. (1400°-1450° C.) |
| | A. C. Davis ** | Lab. experiments | 2552° F. (1400° C.) |
| | | | A reduction of 3 per cent. in CaCO ₃ reduces calcining tem- perature by about 50° C. |

* B.P.C.R.A. stands for British Portland Cement Research Association.

† Private communication.

‡ R. Nacken, *Protokoll der Verhandlungen des Vereins Deutschen Portland-Cement-Fabrikanten* (E.V.), Charlottenburg, 1921, p. 184.

§ E. C. Soper, "Report on a Test of a Portland Cement Plant," *Engineering News*, 1905, liv., Nr. 25, p. 664.

|| H. Kühl, *Protokoll der Verhandlungen des Vereins Deutschen Portland-Cement-Fabrikanten* (E.V.), Charlottenburg, 1922, p. 200.

¶ O. Schott, "Dissertation," 1906, "Clinker Attains."

** A. C. Davis, "Manufacture of Portland Cement" (Falconer, Dublin (1922)).

§ 2 If the maximum temperature that the clinker attains in the furnace is $2,500^{\circ}\text{F}$ ($1,371^{\circ}\text{C}$), and if the hot clinker passing away from the kiln is made to give up its heat to the incoming air, the air cannot be preheated to a higher temperature than $2,500^{\circ}\text{F}$ ($1,371^{\circ}\text{C}$), or it would become hotter than the clinker from which it derives its heat

§ 3 The weight of air required to burn 1 lb of standard coal has previously been shown to be 10.478 lbs (see Chapter X)

It is of interest to ascertain what weight of clinker issuing from the furnace at $2,500^{\circ}\text{F}$ ($1,371^{\circ}\text{C}$) is required in order to raise 10.478 lbs of air to $2,500^{\circ}\text{F}$ ($1,371^{\circ}\text{C}$)

The calculation may be carried out as follows —

The specific heat of Portland cement clinker has been determined very exactly over a wide range of temperature by S. H. Harrison, M.E., and Dr. Walter P. White, of the Geophysical Laboratory,* Washington, U.S.A., who gave the following formula for the mean specific heat of clinker between 0° and $t^{\circ}\text{C}$ —

$$S_m = 0.1754 + 1.3914t \times 10^{-5} - 1.25 \times 10^{-9}t^2 + 4.685 \times 10^{-14}t^3 \quad (1)$$

Now, assuming that the temperature of the clinker is $2,500^{\circ}\text{F}$, or $1,371^{\circ}\text{C}$, we obtain the mean specific heat S_m of the clinker between 0° and $1,371^{\circ}\text{C}$ by inserting $t = 1,371$ in formula (1), and thus obtain

$$S_m = 0.2519 \text{ between } 0^{\circ} \text{ and } 1,371^{\circ}\text{C} \quad (3^{\circ} \text{ to } 2,500^{\circ}\text{F})$$

From our tables of mean specific heats the mean specific heat of air between 32° and $2,500^{\circ}\text{F}$ is 0.2533. Hence, on the assumption that x lbs of the clinker in sinking from $2,500^{\circ}$ to 60°F heats 10.478 lbs of air from 60° to $2,500^{\circ}\text{F}$, we have

$$\begin{array}{ccccccc} \text{Weight of} & \times & \text{Specific Heat} & \times & \text{Range of} & \text{Weight} & \times & \text{Specific} & \times & \text{Range of} \\ \text{Clinker} & & \text{of Clinker} & & \text{Temperature} & \text{of Air} & & \text{Heat of} & & \text{Temperature} \\ & & & & \text{Degree} & & & \text{Air} & & \text{Degree} \\ x & \times & 0.2519 & \times & (2500-60) & = & 10.478 & \times & 0.2533 & \times & (2500-60) \end{array}$$

or

$$\begin{aligned} x &= 10.478 \times \frac{0.2533}{0.2519} \\ &= 10.54 \text{ lbs of clinker} \end{aligned}$$

So that 10.54 lbs of clinker at $2,500^{\circ}\text{F}$ can preheat 10.478 lbs of air from 60° to $2,500^{\circ}\text{F}$

It will be noticed that the mean specific heat of air and clinker may, for practical purposes, be taken as equal

§ 4 In Chapter XIV a table was given (Table I) showing the number of lbs of clinker which is theoretically producible by the combustion of 1 lb of standard coal when the flame temperature attains different values

It is of considerable practical importance in the design of rotary kilns to know to what temperature the incoming air necessary for combustion of the coal can be heated by the outgoing clinker

The calculation may be carried out in the same way as § 3

For example, with a theoretical flame temperature of $4,160^{\circ}\text{F}$ ($2,293^{\circ}\text{C}$), 1 lb of standard coal could produce 10.397 lbs of clinker, and this clinker, issuing at $2,500^{\circ}\text{F}$, could preheat the incoming 10.478 lbs of air necessary for the combustion of the 1 lb of coal to a temperature degree $T^{\circ}\text{F}$, which is determined by the following equation —

* Report on the Specific Heat of Portland Cement Clinker by S. H. Harrison M.E. Comm. tee on Conservation Portland Cement Association of U.S.A. (1933)

$$\begin{array}{ccccccc}
 \text{Weight of Air} & \times & \text{Specific Heat of Air at } 2,500^{\circ} \text{ F.} & \times & \text{Range of Temperature Degree} & = & \text{Weight of Clinker} \times \text{Specific Heat of Clinker} \times \text{Range of Temperature Degree} \\
 10.478 & \times & 0.2533 & \times & (T - 60) & = & 10.397 \times 0.2519 \times (2500 - 60)
 \end{array}$$

or $T^{\circ} = 2,468^{\circ} \text{ F.}$

The next weight of *clinker* on the list of Table I., Chapter XIV., is 10.125 lbs., corresponding to a flame temperature of $4,100^{\circ} \text{ F.}$ ($2,260^{\circ} \text{ C.}$). As there is less clinker now produced per 1 lb. of coal burnt, the 10.478 lbs. of air will not be preheated to quite such a high temperature as before, and we must make an allowance for the variation in the specific heat of the air, which is much less at low temperatures than at high temperatures.

This is done as follows: We take the former temperature of $2,468^{\circ} \text{ F.}$ attained by the air in the preceding case to be the approximate temperature of the air attained in the present case. We then take the mean specific heat of air at $2,468^{\circ} \text{ F.}$ to be the correct value of the mean specific heat to put in our equation for determining T° , the temperature to which the air can be heated by the clinker. We thus obtain the new equation for determining T , thus:—

$$\begin{array}{ccccccc}
 \text{Weight of Air} & \times & \text{Specific Heat of Air at } 2,468^{\circ} \text{ F.} & \times & \text{Range of Temperature Degree} & = & \text{Weight of Clinker} \times \text{Specific Heat of Clinker} \times \text{Range of Temperature Degree} \\
 10.478 & \times & 0.2530 & \times & (T - 60) & = & 10.125 \times 0.2519 \times (2500 - 60)
 \end{array}$$

whence $T = 2,408^{\circ} \text{ F.}$

To aid calculation a general formula may be worked out as follows:—

If S_m be the mean specific heat of the air between T° and 60° F. , and W be the weight of clinker produced, we have, as before

$$10.478 \times S_m \times (T - 60) = W \times 0.2519 \times (2500 - 60),$$

$$\text{or} \quad T^{\circ} = 60 + 58.66 \frac{W}{S_m} \quad . \quad . \quad . \quad . \quad . \quad (2)$$

By substituting in this formula successive values of W from Table I., Chapter XIV., and the corresponding successive values of S_m , corresponding to the various values of T° , the calculation may be easily effected.

§ 5. By proceeding in this way, the table on following pages was calculated, showing the maximum temperature to which the incoming air can be preheated by different weights of outgoing clinker, the clinker being assumed to attain a maximum temperature of $2,500^{\circ} \text{ F.}$ in the furnace.

§ 6. On looking down this table it will be seen that in the rotary kilns consuming 28.2 to 34.3 tons of standard coal per 100 tons clinker, it would be possible to preheat the entering air from 935° to 781° F.

Since at least 15 per cent. of all the air required enters *cold* with the coal dust, the remaining air passing up the clinker shoot could be heated to temperatures about 15 per cent. in excess of these temperatures. The heavy loss of heat from the coolers by radiation, however, would greatly reduce the efficiency of the heat interchange between the air and the clinker, and it may well be doubted whether in many modern rotary kilns the mean temperature of air where it meets the coal dust possesses an average temperature much exceeding 400° F.

There is obviously much room for improvement in the coolers of the ordinary rotary kiln.

The lower the clinker output of the kiln per 1 lb. coal burnt, the lower is the temperature to which the air can be preheated.

§ 7. Several points in this table call for comment. We have seen that the incoming air cannot be heated by the outgoing clinker to a temperature higher

TABLE II

Maximum Temperature to which Incoming Air can be Preheated by the Outgoing Clinker in Kilns Producing Different Yields of Clinker

Assumption — 1 lb standard coal requires 10.478 lbs air Mean specific heat of clinker between 60° and 2,500° F is 0.2519

Yield of Clinker

| (1) Lbs of Clinker Produced per 1 Lb of Standard Coal of 12 600 B Th U s Fired in Kiln W | (2) Tons of Standard Coal Consumed per 100 Tons Clinker Produced $\frac{100}{W}$ | (3) Maximum Temperature in Degrees F to which Incoming Combustion Air can be Preheated by Outgoing Clinker $T = 60^\circ + 58.66 \frac{W}{S_m}$ 10.478 Lbs of Air per 1 Lb of Standard Coal Burnt | |
|--|---|---|------|
| | | ° F | ° C |
| 0 000 | | 60 | 15.6 |
| 0 065 | 1538.5 | 76 | 24 |
| 0 411 | 243.31 | 164 | 73 |
| 0 760 | 131.58 | 257 | 122 |
| 1 111 | 90.00 | 340 | 171 |
| 1 466 | 70.13 | 428 | 220 |
| 1 824 | 54.82 | 516 | 269 |
| 2 185 | 45.77 | 605 | 318 |
| 2 549 | 39.25 | 694 | 368 |
| 2 915 | 34.30 | 781 | 416 |
| 3 283 | 30.46 | 872 | 467 |
| 3 320 | 30.12 | 881 | 472 |
| 3 359 | 29.77 | 891 | 477 |
| 3 395 | 29.45 | 899 | 487 |
| 3 432 | 29.14 | 910 | 488 |
| 3 469 | 28.83 | 917 | 492 |
| 3 507 | 28.51 | 926 | 497 |
| 3 545 | 28.21 | 935 | 502 |
| 3 582 | 27.92 | 944 | 507 |
| 3 619 | 27.63 | 953 | 512 |
| 3 656 | 27.35 | 962 | 517 |
| 3 693 | 27.08 | 970 | 521 |
| 3 731 | 26.80 | 979 | 526 |
| 3 769 | 26.53 | 988 | 531 |
| 3 806 | 26.27 | 997 | 536 |
| 3 843 | 26.02 | 1006 | 541 |
| 3 882 | 25.76 | 1015 | 546 |
| 3 919 | 25.52 | 1024 | 551 |
| 3 956 | 25.28 | 1033 | 556 |
| 3 993 | 25.04 | 1041 | 561 |
| 4 032 | 24.80 | 1047 | 564 |
| 4 413 | 22.66 | 1137 | 614 |
| 4 795 | 20.85 | 1226 | 663 |
| 5 181 | 19.30 | 1315 | 713 |

of the air remains steady at $2,500^{\circ}\text{F}$ in the table, while the clinker produced per 1 lb of coal burnt increases from 10.54 lbs of clinker to as high as 15.732 lbs of clinker, which is the theoretical maximum amount producible

§ 8 This opens up another question Up to the point where the yield of clinker reached 10.54 lbs per 1 lb of standard coal burnt, the coal is supplied with the requisite amount of air—namely, 10.478 lbs—needed for normal combustion, which is preheated to the maximum temperature of $2,500^{\circ}\text{F}$ by the clinker. But when the yield of clinker per 1 lb of coal increases beyond 10.54 lbs, the amount of outpouring clinker is sufficient to preheat *more than* 10.478 lbs of incoming air to $2,500^{\circ}\text{F}$ So that it would in such a case be possible to supply the 1 lb of coal with *more than* the necessary 10.478 lbs of air preheated to $2,500^{\circ}\text{F}$

The exact weight of air which can be thus supplied is shown in the following table, No III

The weight of air is calculated from the equation

$$\begin{aligned} & \text{weight of air} \times \text{specific heat of air} \times \text{rise of temperature degree of air} \\ & = \text{weight of clinker} \times \text{specific heat of clinker} \times \text{fall in temperature of clinker} \end{aligned}$$

Taking the mean specific heat of the air between 60° and $2,500^{\circ}\text{F}$ as 0.2533 and that of clinker as 0.2519 we get, if W be the required weight of air—

$$W \times 0.2533 \times (2500 - 60) = \text{weight of clinker} \times 0.2519 \times (2500 - 60),$$

which reduces to

$$\text{weight of air} = 0.9945 \times \text{weight of clinker}$$

TABLE III

Showing Weight of Air in Lbs which can be Preheated to $2,500^{\circ}\text{F}$ by a Given Weight of Clinker

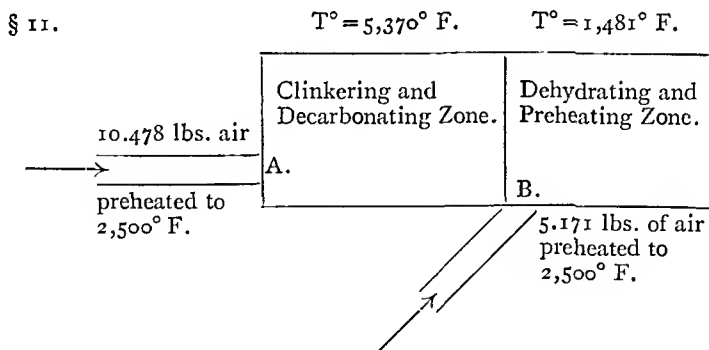
| Tons of Standard Coal Consumed per 100 Tons of Clinker | Lbs of Clinker Produced per 1 Lb of Standard Coal Burnt of 12 600 B Th U s per Lb W | Weight of Air which can be Preheated to $2,500^{\circ}\text{F}$ by the Clinker 0.9945W | Weight of Air in Excess of that Needed for Normal Combustion which can be Preheated to $2,500^{\circ}\text{F}$ by the Clinker 0.9945W 10.478 |
|--|---|--|--|
| | | Lbs | Lbs |
| 9.49 | 10.540 | 10.478 | 0.000 |
| 9.37 | 10.675 | 10.620 | 0.142 |
| 9.13 | 10.954 | 10.896 | 0.418 |
| 8.90 | 11.239 | 11.178 | 0.700 |
| 8.68 | 11.522 | 11.461 | 0.983 |
| 8.47 | 11.806 | 11.744 | 1.266 |
| 8.27 | 12.091 | 12.028 | 1.550 |
| 8.07 | 12.386 | 12.322 | 1.844 |
| 7.89 | 12.675 | 12.607 | 2.129 |
| 7.71 | 12.972 | 12.902 | 2.424 |
| 7.53 | 13.271 | 13.200 | 2.722 |
| 7.37 | 13.575 | 13.503 | 3.025 |
| 7.21 | 13.873 | 13.796 | 3.318 |
| 7.05 | 14.177 | 14.100 | 3.622 |
| 6.90 | 14.485 | 14.405 | 3.927 |
| 6.76 | 14.793 | 14.713 | 4.235 |
| 6.62 | 15.102 | 15.028 | 4.550 |
| 6.49 | 15.414 | 15.330 | 4.852 |
| 6.36 | 15.732 | 15.649 | 5.171 |

§ 9. Hence it appears that if a kiln were constructed so as to yield anything over 10.54 lbs. of clinker per 1 lb. of coal burnt (or 100 tons clinker per 9.49 tons of standard coal of 12,600 B.Th.U.'s per lb.), it would be possible to feed the coal with air in excess of that needed for normal combustion, all this air being preheated to the clinkering temperature of 2,500° F. (1,371° C.) by the outgoing clinker; and when the theoretical limit of clinker production was reached (viz., 100 tons clinker for 6.36 tons of standard coal, or 15.732 lbs. of clinker per 1 lb. of standard coal consumed), it would be possible under these circumstances to supply no less than 5.171 lbs. of excess air thus heated to each 1 lb. of coal consumed—a very large amount of hot air, for, since 1 lb. of coal only requires 10.478 lbs. of air, the excess air would be $\frac{5.171}{10.478} \times 100 = 49.3$ per cent. excess air over that needed for normal combustion.

§ 10. The question now arises: Is there any advantage to be derived in thus feeding the kiln with great excess of air preheated to 2,500° F. (1,371° C.) by the outgoing clinker?

The answer is: Yes—if we only supply the normal amount of air, no increased yield of clinker will result whether this hot excess air be introduced either with the coal through the clinkering zone, or whether it be introduced at a later stage than the clinkering zone, and be used for preheating the slurry as it enters the kiln.

But this excess of hot air would certainly have to be introduced into the kiln for heat economy, otherwise (1) it would not be possible to recover all the heat from the clinker and put it back into the kiln (this is the calculated weight of air which just achieves this purpose); (2) there would not be sufficient total heat supplied to the lower part of the kiln to carry out all the necessary operations of preheating the slurry and dehydrating the clay so as to yield 15.732 lbs. of clinker per 1 lb. of coal burnt. A heat balance in detail has been worked out which shows that this is the case:—



Suppose that the kiln is producing the maximum amount possible of clinker (see Chapter XIV.), namely, 15.732 lbs. of clinker per 1 lb. of standard coal burnt (of 12,600 B.Th.U.'s per lb.), and suppose that *all* the heat escaping with the clinker is restored to the furnace by 15.649 lbs. of entering air, which is thus preheated to 2,500° F., as shown in Table II.

For convenience in following what will occur, we will allow the preheated air to enter in two portions, viz., 10.478 lbs. at A—the exact amount necessary for burning 1 lb. of coal—and the excess 5.171 lbs. of hot air at B.

The 10.478 lbs. of air entering at A, and there uniting with 1 lb. of standard coal, will produce a theoretical flame temperature of 5,370° F., as shown in Chapter XII., and the 11.278 lbs. of furnace gas passing down the clinkering and

decarbonating zone and issuing at $1,481^{\circ}\text{F}$ will produce in this zone (as shown in Chapter XIV) 15 732 lbs of clinker as the result of decomposing 24 547 lbs of dry slurry. If now the hot gases evolved in the formation of this weight of clinker be allowed to proceed down the kiln into the dehydrating and preheating zone, it can be easily calculated that they do not contain enough heat to do the necessary work of preheating and dehydrating the slurry before it enters the decarbonating zone. Thus, if we allow the gases to escape from the kiln at 60°F —after having given up all their heat to the raw material—we will find

- (1) That the B Th U's from 11 278 lbs of the gases of combustion sinking from $1,481^{\circ}$ to 60°F are 4,628
- (2) That the B Th U's from the CO_2 from the slurry sinking from $1,481^{\circ}$ to 60°F are 2,925
- (3) That the B Th U's liberated from the water expelled from the kaolin are 699.5
- (4) That the B Th U's liberated from the water in the silica are 233.6

Hence the total available heat from the gases is

$$4628 + 2925 + 699.5 + 233.6 = 8486.9 \text{ B Th U's}$$

But, as shown in Chapter IX, in order to preheat 24 547 lbs of slurry from 60° to $1,481^{\circ}\text{F}$, we will require 10 265 B Th U's

Hence there is not enough heat in the furnace gases to do this, and there is missing

$$10265 - 8487 = 1,778 \text{ B Th U's}$$

So that what would happen unless this extra amount of heat was supplied to the lower part of the kiln, is that the cold undehydrated and unpreheated raw slurry would pass into the decarbonating zone, so that the heat used for expelling the CO_2 from the calcium carbonate would now be utilised in preheating the slurry. So that less calcium carbonate would be now decomposed, and this would reduce the clinker formed until finally the missing heat would be supplied to the lower zone at the expense of the upper zone, until equilibrium was once more obtained. A decrease in clinker output, therefore, could only be avoided by allowing the extra 5 173 lbs of hot air (preheated to $2,500^{\circ}\text{F}$) to enter the preheating zone at the point B. This could supply 1,786 B Th U's in passing down the kiln, and so supply the missing amount of heat and allow a clinker production of 15 732 lbs per 1 lb of coal to proceed continuously in the upper zone.

In other words, we have proved from the heat balance that under these special conditions the kiln must receive more hot air than corresponds to normal combustion (10 478 lbs of air per 1 lb of coal) if the output of the kiln is to be kept up.

Of course, the net effect would be the same if the excess of hot air was allowed to enter with the coal at A in the upper part of the kiln, instead of at B. An initial increased output of clinker from the upper part of the kiln would be speedily brought down to the above figure on account of the gases not containing sufficient heat to effect the necessary preheating and dehydrating operations in the lower part of the kiln.

§ 12 Another important point which arises in connection with matter discussed in §§ 10 and 11 is this. Would it be possible to increase the output of clinker by supplying a kiln with air in excess of that normally needed for combustion (112, 10 478 lbs of air to 1 lb of standard coal), this air preheated by the outgoing clinker in such a way that all the heat contained in the outpouring clinker is restored to the furnace by the air?

It is proved below that the answer to this question is No, except in the very special case (discussed in §§ 10 and 11 above), where the yield of clinker per 1 lb of coal burnt is exceeding 10.54 lbs.

CHAPTER XVII

PROOF THAT THE CLINKER OUTPUT PER 1 LB.
OF COAL BURNT IS GREATEST WHEN THE
FLAME TEMPERATURE IS HIGHEST, AND
THAT THE OUTPUT OF CLINKER MUST
DIMINISH WHEN AIR IN EXCESS OF 10.478
LBS. PER 1 LB. OF STANDARD COAL IS
SUPPLIED TO THE KILN

§ 1. IN Chapter XIV. we showed that the weight of clinker produced depends upon the quantity of heat Q contained in the gases of combustion evolved from the combustion of 1 lb. of coal, and that in general the higher the initial flame temperature, the greater the clinker output.

In the whole chapter, however, it is assumed that the air supplied to the kiln is that necessary for normal combustion—viz., 10.478 lbs. of air per 1 lb. of standard coal consumed.

§ 2. It is the purpose of this chapter to prove that any quantity of air supplied beyond 10.478 lbs. per 1 lb. of coal must necessarily diminish the clinker output except in one particular case, which so far has not been attained in practice and which will be discussed below.

§ 3. *To prove that the maximum output of clinker per 1 lb. of coal burnt is attained when 10.478 lbs. of air are supplied per 1 lb. of standard coal, and that any excess of air beyond this lowers the clinker output per 1 lb. of coal burnt.*

Assume the clinkering temperature of the material to be $2,500^{\circ}$ F. and the temperature of the atmosphere to be 60° F. Let W_1 lbs. of clinker be produced per 1 lb. of standard coal consumed when the air supply is the normal one of 10.478 lbs. per 1 lb. of coal burnt. The weight of combustion gases produced is 11.278 lbs. per 1 lb. of coal (see Chapter X.). Let t_1° be the temperature of the 10.478 lbs. of entering air when it meets the coal. It is assumed that the W lbs. of outgoing clinker has given up all its heat to the incoming 10.478 lbs. of air, which would be the case in a perfect kiln.

Hence t_1° is always less than $2,500^{\circ}$ F.

Let T_1 be the flame temperature attained when the 10.478 lbs. of air at t_1° F. unite with 1 lb. of coal.

Let S_1 be the mean specific heat of the furnace gases between T_1 and $1,481^{\circ}$ F.—the temperature at which the CaCO_3 decomposes (see Chapter VII.).

Then the B.Th.U.'s evolved when 11.278 lbs. of combustion gas sink from T_1° to $1,481^{\circ}$ F. are

$$Q_1 = 11.278 \times S_1 \times (T_1 - 1481) \quad . \quad . \quad . \quad (1)$$

also the weight of clinker W_1 is given by (see Chapter XIV)

$$W_1 = \frac{Q_1}{918.6} \quad (2)$$

Let S_2 be the mean specific heat of the 10 478 lbs of entering air between 60° and $t_1^\circ \text{F}$ and S_3 the mean specific heat of the clinker between 2500° and 60°F

Then the temperature t_1 of the entering air is given by

$$10\,478 \times S_2 \times (t_1 - 60) = W_1 \times S_3 \times (2500 - 60) \quad (3)$$

$$\begin{aligned} &= \frac{Q_1}{918.6} \times S_3 \times 2440 \\ &= 2\,656 Q_1 \times S_3 \end{aligned}$$

whence
$$t_1 = 0.2535 Q_1 \times \frac{S_3}{S_2} + 60 \quad (4)$$

Now let an additional X lbs of air be supplied to the kiln. This air will come in cold because all the heat has been already abstracted from the clinker by the

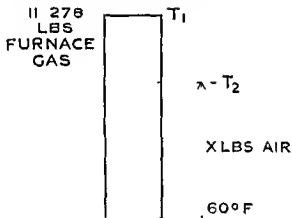


FIG. 1

10 478 lbs (so that t_1 is kept unchanged while the other X lbs of air enters at 60°F). Let the new flame temperature be T_2 and the new mean specific heat of the furnace gas between T_1 and T_2 be S_4 and let the mean specific heat of air between 60° and T_2 be S_5

Then we have 11 278 lbs of furnace gas sinking from T_1 to T_2 will heat x lbs of air from 60° to T_2

$$\text{So that } 11\,278 \times S_4 \times (T_1 - T_2) = X \times S_5 \times (T_2 - 60)$$

$$\text{or } T_2(X \times S_5 + 11\,278 \times S_4) = T_1 \times 11\,278 \times S_4 + 60X \times S_5$$

$$\text{or } T_2 = \frac{11\,278 \times S_4 \times T_1 + 60X \times S_5}{X \times S_5 + 11\,278 \times S_4} \quad (5)$$

We have now 11 278 lbs of combustion gas at $T_2^\circ \text{F}$ and X lbs of air at $T_2^\circ \text{F}$ and this mass of gas sinking from T_2° to 1481°F will liberate an amount of heat Q_2 which will be available for forming the new weight of clinker—

$$W_2 = \frac{Q_2}{918.6} \quad (6)$$

Then we must show that the new amount of heat Q_2 evolved is *less than* the old amount Q_1 , in order that W_2 may be less than W_1 .

$$\text{Now} \quad Q_2 = 11.278 \times S_6 \times (T_2 - 1481) + X \times S_7 (T_2 - 1481),$$

where S_6 and S_7 is the mean specific heat of the combustion gases and of air between T_2° and $1,481^\circ \text{ F.}$,

$$\text{or} \quad Q_2 = T_2 (11.278 \times S_6 + X \times S_7) - 1481 (11.278 S_6 + X \cdot S_7).$$

Now substitute the value of T_2 from (4) and we see

$$\begin{aligned} Q_2 = & \left(\frac{11.278 S_6 + X \times S_7}{11.278 S_4 + X \cdot S_5} \right) T_1 \times 11.278 \times S_4 \\ & + 60X \times S_5 \left(\frac{11.278 S_6 + X \cdot S_7}{11.278 S_4 + X \cdot S_5} \right) \\ & - 1481 (11.278 S_6 + X \cdot S_7) \quad . \quad . \quad . \quad . \quad (7) \end{aligned}$$

and from (1)

$$Q_1 = 11.278 S_1 T_1 - 1481 (11.278 S_1) \quad . \quad . \quad . \quad . \quad (8)$$

$$\begin{aligned} \text{whence} \quad Q_1 - Q_2 = & T_1 \left(11.278 S_1 - \frac{11.278 S_6 + X \cdot S_7}{11.278 S_4 + X \cdot S_5} \times 11.278 \times S_4 \right) \\ & - 1481 (11.278 S_1 - 11.278 S_6) \\ & + 1481 X \cdot S_7 - 60X \cdot S_5 \frac{11.278 S_6 + X \cdot S_7}{11.278 S_4 + X \cdot S_5}. \end{aligned}$$

$S_1, S_2, S_3, S_4, S_5, S_6$, and S_7 —the specific heats of the air, combustion gases, and clinker at temperatures above $1,481^\circ \text{ F.}$ —are all very nearly equal.

Whence, putting $S_1 = S_2 = S_3 = S_4 = S_5 = S_6 = S_7 = 0.3$, we get

$$Q_1 - Q_2 = 1421 X \cdot S_1.$$

Now S_1 , the specific heat, is essentially a positive number, and X , the excess air, added to the 10.478 lbs. can only vary from 0 to a positive number.

Hence $Q_1 - Q_2 > \text{a positive number,}$

or $Q_1 > Q_2.$

Consequently, Q_2 is always less than Q_1 when X is greater than zero.

$$\text{But} \quad W_2 = \frac{Q_2}{918.6} \quad . \quad . \quad . \quad . \quad (9)$$

$$\text{and} \quad W_1 = \frac{Q_1}{918.6} \quad . \quad . \quad . \quad . \quad (10)$$

Hence $W_1 > W_2$ whenever X is greater than 0.

Substituting in (7)

$$\therefore W_1 - W_2 = \frac{1421}{9186} \times X \times S_1 \quad . \quad . \quad . \quad . \quad (11)$$

Now, taking S_1 as approximately 0.3 for high temperatures, we get

$$W_1 - W_2 = \frac{1421 \times 0.3}{918.6} \lambda$$

$$W_1 - W_2 = 0.46\lambda \text{ approximately} \quad (12)$$

or

$$W_2 - W_1 = 0.46X \quad (12a)$$

which shows that in general the greater is λ —the excess air over 10.478 lbs—the greater is the loss in clinker output. This proves the proposition.

Of course, (12) is only an approximate formula, but is useful in approximately calculating the diminution of output in clinker caused by increasing the air supply. For ordinary kilns—when the flame temperature is only about 2,600° F— S_1 may be taken as 0.252,

whence

$$W_1 - W_2 = 0.4X,$$

or

$$W_2 - W_1 = 0.4X$$

§ 4 Hence, so long as t_1 —the temperature of the entering air heated by the clinker—is below the clinkering temperature 2,500° F, the clinker output is greatest when $\lambda = 0$, i.e., when only 10.478 lbs of air are supplied per 1 lb of standard coal.

But under these conditions it was proved in Chapter XIV that the clinker output is greater the higher the flame temperature. So that we have arrived at a conclusion of vital importance in kiln design, namely—

The higher the flame temperature attained in the clinkering zone, the greater the possible output of clinker per 1 lb of coal burnt.

By no possible manipulation (such as introducing excess of air) is it possible to reduce the flame temperature in the clinkering zone without simultaneously reducing the clinker output per 1 lb of coal burnt.

A greater weight of colder furnace gas will never prove so efficient in producing clinker as a smaller weight of very hot gas.

§ 5 On account of the very great practical importance of the above principle we will work out in detail some examples showing the decrease in clinker output which follows an increase in the supply of air beyond 10.478 lbs per 1 lb of coal burnt.

Example—In a perfect kiln using 10.478 lbs of air per 1 lb of coal and producing 9.311 lbs clinker per 1 lb of standard coal fired, the temperature of the entering air (preheated by the clinker so as to abstract all its heat) is 2,236° F and the flame temperature is 3,916° F.

Trace the effect on the clinker output if the air supply is increased from 10.478 to 13.478 lbs per 1 lb of coal.

Here $X = 3$ lbs (= weight of air in excess of 10.478 lbs.)

$$W_1 = 9.311$$

Hence from equation (12)

$$\begin{aligned} W_2 &= W_1 - 0.46X \\ &= 9.311 - 0.46 \times 3 \\ &= 9.311 - 1.380 = 7.931 \text{ lbs} \end{aligned}$$

So that the effect of increasing the air supply by 3 lbs per 1 lb of air would be to diminish the clinker output from 9.3 to 7.9 lbs.

Let us now check this result by carrying the calculation through from first principles. The flame temperature when only 10.478 lbs of gas are present is 3,916° F. The B.T.U.'s evolved when 13.478 lbs of combustion gas sink from

3,916° to 1,481° F. is 8,553 B.Th.U.'s, corresponding to the clinker weight of 9.311 lbs. per 1 lb. of coal burnt.

On now adding 3 lbs. of cold air to the entering air (making 13,478 lbs. of air in all) we must have 11.278 lbs. of furnace gas sinking from 3,916° F. to the new flame temperature T° , heating 3 lbs. of air from 60° to T° .

Taking the mean specific heat of air between 60° and 3,000° F. as 0.2583, and taking the mean specific heat of the furnace gas between 60° and 3,000° F. as 0.3053, then we have

$$11.278 \times 0.3053 \times (3916 - T) = 3 \times 0.2583 \times (T - 60),$$

or $T = 3,208^\circ$ F. as the new flame temperature of the diluted mass of gas.

The mean specific heat of the combustion gas between 1,481° and 3,208° F. is 0.3018 F., and that of air under the same circumstances is 0.275.

Then we have 11.278 lbs. of combustion gas at 3,208° F. + 3.0 lbs. of air at 3,208° F. sinking to 1,481° F.—the temperature at which the CaCO_3 decomposes—will evolve

$$11.278 \times 0.3013 \times 1727 + 3 \times 0.275 \times 1727 = 7,293 \text{ B.Th.U.'s},$$

which are available for clinker formation under the new circumstances.

So that the new amount of clinker formed will now be $\frac{7293}{918.6} = 7.93$ lbs.

In other words, the addition of 3 lbs. of air to the 10.478 lbs. entering the kiln per 1 lb. of coal burnt has lowered the clinker output per 1 lb. of coal fired from 9.311 to 7.93 lbs. This thus checks our former result.

§ 6. *Exceptional case when the temperature of the incoming air is preheated to 2,500° F. (the clinkering temperature) by the clinker, and when the kiln is producing over 10.54 lbs. of clinker per 1 lb. of coal burnt.*

In Chapter XVI., §§ 7 and 8, it was shown that, if a kiln was constructed so as to yield anything between 10.54 and 15.732 lbs. of clinker per 1 lb. of coal burnt (or 100 tons clinker per 9.49 to 6.36 tons of standard coal of 12,600 B.Th.U.'s per 1 lb.), it would be possible to feed the coal with air in excess of that needed for normal combustion (viz., 10.478 lbs. air per 1 lb. of coal), all this air being preheated by the clinker to 2,500° F. (1,371° C.) (the clinkering temperature) by the outgoing clinker, and when the theoretical limit of clinker production was reached (viz., 100 tons of clinker for 6.36 tons of standard coal consumed, or 15.732 lbs. of clinker per 1 lb. of standard coal consumed), it would then be possible under these circumstances to supply no less than 5.171 lbs. of excess air thus heated to each 1 lb. of coal consumed—a very large amount of hot air, for, since 1 lb. of coal only requires 10.478 lbs. of air, the excess air would be

$$\frac{5.171}{10.478} \times 100 = 49.3 \text{ per cent.},$$

i.e., nearly 50 per cent. in excess of that needed for normal combustion. The reader should consult §§ 5, 6, and 7 of Chapter XVI.

§ 7. It was shown in Chapter XVI. that under the above circumstances, if the supply of hot air to the kiln was restricted to 10.478 lbs. per 1 lb. of coal, the kiln could only produce 10.54 lbs. of clinker. But if the supply of hot air was gradually increased from 10.478 to 15.649 lbs., the output of clinker would likewise increase from 10.54 to 15.732 lbs. per 1 lb. of coal burnt.

Hence this fact seems at first sight to contradict the conclusion of § 3—that if the air supply be increased beyond 10.478 lbs. per 1 lb. coal, the output must diminish.

§ 8 A closer examination of the theory, however, will show that this case is also included in the general one, consistent with the restriction of the value of t_1 (the temperature of the incoming air as preheated by the outgoing clinker to the point where it abstracts all the heat from the clinker) to $2,500^\circ \text{F}$, since no matter how much clinker is produced, the clinker cannot preheat the air to above $2,500^\circ \text{F}$, as this would mean that the air can be made hotter than the outgoing heating material, which is contrary to the second law of thermodynamics.

Consider now equation (3) of § 3, on which the above theory was built up

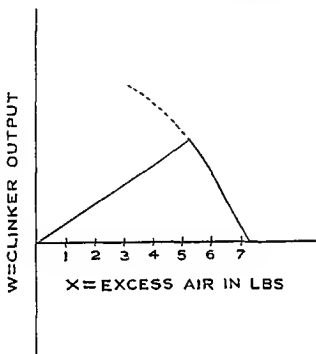


FIG 2

We have the following relationship for the temperature t to which the given weight of air can be preheated by the outgoing clinker —

$$\left\{ \begin{array}{l} \text{Weight of entering air} \\ \text{per 1 lb of coal} \end{array} \right\} \times \left\{ \begin{array}{l} \text{Mean specific heat of air} \\ \text{between } 60^\circ \text{ and } t^\circ \text{ F} \end{array} \right\} \times \left\{ \begin{array}{l} \text{Rise in temperature} \\ \text{of air} \end{array} \right\} \\ = \left\{ \begin{array}{l} \text{Weight of outgoing} \\ \text{clinker per 1 lb} \\ \text{of coal} \end{array} \right\} \times \left\{ \begin{array}{l} \text{Mean specific heat of} \\ \text{clinker between} \\ 60^\circ \text{ and } 2,500^\circ \text{ F} \end{array} \right\} \times \left\{ \begin{array}{l} \text{Fall in temperature} \\ \text{of clinker} \end{array} \right\},$$

or, if λ be the weight of air in lbs in excess of the normal weight 10.478 required for the combustion of 1 lb of coal, and W be the weight of clinker produced per 1 lb of coal burnt—

$$(10.478 + X) \times \left\{ \begin{array}{l} \text{Mean specific heat} \\ \text{of air between} \\ 60^\circ \text{ and } t^\circ \end{array} \right\} \times (t - 60) = W \times 0.2519 \times (2500 - 60) \quad (1)$$

Now when the weight W of clinker reaches 10.54 lbs per 1 lb of coal burnt, we have seen that $t = 2,500^\circ \text{F}$ (see Chapter XVI, § 3), and specific heat of air between 60° and $2,500^\circ \text{F}$ becomes 0.2533

So that (1) reduces itself to

$$(10.478 + X) \times 0.2533 \times (2500 - 60) = W \times 0.2519 \times (2500 - 60),$$

or the factor $(2500 - 60)$ cancels out on both sides, and we get, *so long as* $t = 2,500^\circ \text{F.}$

$$10.478 + X = 0.9945W \quad (2)$$

This shows that, *so long as* $t = 2,500^\circ \text{F.}$, the weight W of the clinker will steadily increase with X —the weight of excess air. For example, if

$$X = 0, \quad W = \frac{10.478}{0.9954} = 10.54 \text{ lbs. of clinker.}$$

$$X = 1, \quad W = \frac{11.478}{0.9954} = 11.5 \quad \text{,,} \quad \text{,,}$$

$$X = 2, \quad W = \frac{12.478}{0.9954} = 12.5 \quad \text{,,} \quad \text{,,}$$

$$X = 3, \quad W = \frac{13.478}{0.9954} = 13.5 \quad \text{,,} \quad \text{,,}$$

$$X = 4, \quad W = \frac{14.478}{0.9954} = 14.5 \quad \text{,,} \quad \text{,,}$$

$$X = 5.171, \quad W = \frac{15.649}{0.9954} = 15.7 \quad \text{,,} \quad \text{,,}$$

(See Table III., Chapter XVI.)

But equation (2) is subjected to another restriction as well. W cannot exceed 15.732 lbs. of clinker per 1 lb. of coal burnt (see Chapter XIV.) on account of the heat necessary to form the clinker and decompose the CaCO_3 .

Hence the excess air X can only range from 0 to 5.171 lbs., and within these limits the *yield of clinker* W per 1 lb. of coal burnt will increase from 10.54 to 15.73 lbs. per 1 lb. of coal burnt.

If X increases beyond 5.171 lbs., then the former law will come into operation and the clinker output will again begin to decrease with X .

It will be seen, therefore, that the new relationship holds only for the very special case for the brief range when the temperature of the incoming air is heated to the clinkering temperature $2,500^\circ \text{F.}$ by the outgoing clinker, and is a direct consequence of the second law of thermodynamics—viz., that the air cannot be heated to a higher temperature than $2,500^\circ \text{F.}$ by clinker issuing at $2,500^\circ \text{F.}$

§ 9. From the preceding article it is clear that when the air supply increases beyond certain limits, the value of Q diminishes. It will be remembered that Q is the amount of heat available for clinker formation contained in the furnace gases above $1,481^\circ \text{F.}$ It is, therefore, of interest to examine the values of X for which Q will become zero—i.e., when clinker will cease to be formed when the excess air exceeds certain limits.

The calculations will be much simplified by assuming that the specific heats of the clinker, hot air, and furnace gases are all equal to each other. This is sufficiently correct not to give misleading results. We will also take this specific heat to be equal to 0.30, as in most cases the gases are at a high temperature.

Let $t^\circ \text{F.}$ = the initial temperature of the air at the point where combustion commences.

T° = the flame temperature when the air is at t° .

X = weight of air added (in lbs.) in excess of that needed for normal combustion.

S = mean specific heat of gases between t° and $T^\circ \text{F.} = 0.30$.

Then 11.78 lbs of combustion gas + λ lbs of excess air at t° F are raised to T° F by 17600 B Th U's developed by the combustion of 1 lb of standard coal

Hence

$$\begin{array}{ccccccc} \text{Weight of Gas} & \times & \text{Specific Heat} & \times & \text{Rise of Temperature} & = & \text{B Th U's Available for Heating} \\ (11.78 + \lambda) & \times & 0.3 & \times & (T - t) & = & 17,600, \end{array}$$

$$\text{or} \quad T = \frac{17,600}{(11.78 + \lambda) \times 0.3} + t \quad (1)$$

The quantity of heat Q available for clinker formation is obtained by (11.78 + λ) lbs of hot gas sinking from T° to $1,481^\circ$ F (the temperature of decomposition of the CaCO_3)

$$\text{Hence} \quad Q = (11.78 + \lambda) \times 0.3 \times (T - 1481) \text{ B Th U's} \quad (2)$$

Now, substituting the value of T from (1), we get

$$Q = 7589.19 + 3.3834\lambda + 0.3\lambda t - 444.3\lambda \quad (3)$$

Now take the mean specific heats of the air and clinker between 60° and $2,500^\circ$ F as being equal to each other and equal to 0.25 which is nearly true (see § 3 Chapter XVI)

Then

$$\begin{array}{ccccccc} \text{Weight of Clinker} & \times & \text{Specific Heat of Clinker} & \times & \text{Fall of Temperature} & = & \text{Weight of Entering Air} \times \text{Specific Heat of Air} \times \text{Rise of Temperature of Air} \\ \text{or } \frac{Q}{918.6} & \times & 0.25 & \times & (2500 - 60) & = & (10.478 + \lambda) \times 0.25 \times (t - 60), \end{array}$$

which gives

$$t = \frac{2,656Q}{10.478 + \lambda} + 60 \quad (4)$$

Now, substituting this value of t in (3) we get

$$Q = \frac{81647 + 3325\lambda - 426\lambda^2}{1.492 + 0.2032\lambda} \quad (5)$$

From (5) we see that there are two values of λ which reduce Q to zero, these values being the roots of the quadratic equation

$$426\lambda^2 - 3325\lambda - 81647 = 0,$$

whence $\lambda = +18.3$ lbs nearly or $\lambda = -10.5$ lbs nearly

In other words, if the normal air supply of 10.478—say, 10.5 lbs—of air per 1 lb of coal is increased by 18.3 lbs, making 10.5 + 18.3 = 28.8 lbs of air supplied per 1 lb of coal, we would so reduce our flame temperature to nearly $1,481^\circ$ F and so render it impossible to form clinker (see Chapter XIV) as this is the temperature of decomposition of the CaCO_3 .

On the other hand, if we diminished our normal air supply by 10.5 lbs ($t.e.$, from 10.5 lbs per 1 lb of air to 10.5 - 10.5 = 0 lbs), we would also stop clinker formation because the air supply would fail.

§ 10 Let us examine equation (5) in § 9 more carefully

$$\text{Here} \quad Q = \frac{81647 + 3325\lambda - 426\lambda^2}{1.492 + 0.2032\lambda} \quad (5)$$

Put $X=0$. Then $Q=54,723$ B.Th.U.'s. What does this mean? On arriving at (5) it will be recollected that the values of t were left absolutely unrestricted, whereas in practice the values of t are restricted by the laws of thermodynamics to under $2,500^{\circ}$ F. The equation (5) simply asserts that if the second law of thermodynamics did not hold, and the heat from the outgoing clinker could be employed in heating the incoming air *above* $2,500^{\circ}$ F., it would be possible to so preheat the furnace gases as to make them yield $54,723$ B.Th.U.'s per 1 lb. of coal burnt, and so produce $54,723 \div 918.6 = 59.6$ lbs. of clinker per 1 lb. of coal burnt. By putting in various other values of X we could obtain how much clinker can be produced under these circumstances by different weights of excess air. As these cases are impossible of realisation under any natural conditions, they will not be further discussed here.

§ 11. The subject, however, has a practical aspect, in that it tells us what *cannot* be done in kiln design, and what are possible measures for increasing output.

Consider, for example, the following very important point for kiln design:—As proved above, air entering a kiln and preheated to $2,500^{\circ}$ F. by the issuing clinker cannot produce any more clinker than 15.732 lbs., even under the most favourable circumstances. Otherwise we could increase indefinitely the output of a kiln by first supplying it with a little more air preheated to $2,500^{\circ}$ F. by the clinker. This will produce by supposition a little more clinker, which could in its turn be utilised to preheat a little more air to $2,500^{\circ}$ F., which again on introduction into the furnace would produce more clinker, and so on indefinitely. So that at first sight it might appear that it would be possible to increase *indefinitely* the amount of clinker per 1 lb. of coal consumed by supplying more and more hot air. This, we have proved above, is impossible. What does happen is that the increase of clinker with increased supply of hot air (heated by the outgoing clinker) tends to a limit which cannot be passed, and when the amount of air is increased beyond these limits, the clinker output begins to rapidly diminish. All devices, therefore, for increasing the efficiency of a kiln based on such false premises must fail.

Conclusion

§ 12. The vital importance of the preceding thermodynamical proofs lies in this:—

A high flame temperature is essential for increased kiln economy per 1 lb. of coal burnt.

In the modern rotary kiln the flame temperature is inordinately low at $2,600^{\circ}$ F., and owing to this fact the rotary kiln is extremely inefficient. Unless the flame temperatures can be increased, *the limits set to kiln economy are narrow.*

Consequently, the attention of kiln designers should be directed towards supplying as much high-grade heat as possible into *the upper part of the kiln*—viz., the clinkering and decarbonating zones. Low-grade heat (such as is used in expelling water) is useless for increasing output of clinker, except in so far as it relieves the high-grade heat of certain work that it should never be allowed to perform, such as the expulsion of water from the slurry.

The same principle is evident in many familiar processes. For example, water boils at 212° F., and if we wish to produce steam at any perceptible pressure, the B.Th.U.'s must enter the water from gases heated *above* 212° F. (100° C.).

A million B.Th.U.'s contained in material below 212° F. are useless for raising steam at pressure greater than atmosphere. All this low-grade heat can do is to *preheat the water to 212° F. and so take the burden off the higher-grade heat for this purpose.*

Similarly, in *lime-burning* the calcium carbonate could be heated for years to,

sav, $1,000^{\circ}\text{F}$ without producing 1 lb of lime Unless the heat is supplied above $1,481^{\circ}\text{F}$ (805°C), lime will not be produced, and the measure of the amount of lime producible is measured by the number of B Th U 's available in the heating gases before they cool to $1,481^{\circ}\text{F}$ (805°C)

Practical Difficulties—There are however, very serious practical difficulties to contend with in increasing the flame temperature in the upper part of the kiln

In the first place, the brickwork will only stand a fairly narrow limit of temperature, so that widely increasing the temperature will cause the brickwork to fail Also the firing of coal dust with very hot air causes difficulties owing to the rapid combustion

These difficulties can be got over in several ways—

- (1) By increasing the refractory nature of the brick
- (2) By scientifically designing the shape of the clinkering zone so as to utilise the known principles of radiation and conduction

CHAPTER XVIII

WEIGHTS OF DRY SLURRY USED AND WEIGHTS OF CARBON DIOXIDE AND MOISTURE EXPELLED THEREFROM IN PRODUCING DIFFERENT WEIGHTS OF CLINKER

IN Chapter XIV. we calculated the weight of clinker producible per 1 lb. of standard coal fired in the upper part of the kiln using different flame temperatures. In order to calculate the exit temperatures and of the resulting combustion gases and the heat balance of the kiln it is necessary to calculate the amount of dry slurry, expelled CO_2 , water, etc., appertaining to the various amounts of clinker produced. This we will now proceed to do from the data given in Chapter IX.

1. Calculation of Weight of Dry Slurry which Produces to a Given Weight of Clinker.—In Chapter IX. it was shown that 100 lbs. of clinker were produced from 156.03 lbs. of dry slurry.

So that x lbs. of clinker are produced by $1.5603x$ lbs. of slurry.

In Chapter XIV. a table was given (Table I.) showing the number of lbs. of clinker theoretically producible per 1 lb. of standard coal fired in the kiln. By multiplying the number of lbs. of clinker shown therein by the factor 1.5603 we obtain the weight of *slurry* from which it was derived, as shown in column (3) of the annexed Table I.

2. Calculation of the Weight of Carbon Dioxide Gas Evolved from the Slurry when Producing a Given Weight of Clinker.—In Chapter IX. it was shown that in the production of 100 lbs. of clinker 52.38 lbs. of CO_2 were expelled from the slurry.

So that in the production of x lbs. of clinker there were expelled 0.5238 lbs. of CO_2 .

In Chapter XIV. a table was given (Table I.) showing the number of lbs. of clinker theoretically producible per 1 lb. of standard coal fired in the kiln. By multiplying this weight of clinker by the factor 0.5238 we derive column (4) of the annexed Table I., which shows the weight of carbon dioxide evolved from the slurry when this weight of clinker is formed in the kiln.

3. Calculation of the Weight of Combined Water Expelled from the Kaolin in the Dry Slurry during the Production of a Given Weight of Clinker.—From the data given in Chapter IX. it is shown that to every 100 lbs. of clinker produced 2.57 lbs. of combined water are expelled from the kaolin in the slurry. Hence the weight of water expelled from the kaolin during the formation of x lbs. of clinker is formed by multiplying x by the factor 0.0257. In this way is derived column (5) of the following table.

4. Calculation of the Weight of Combined Water Expelled from the Hydrated Silica in the Dry Slurry during the Production of a Given Weight of Clinker.—In Chapter IX. it was shown that in the production of 100 lbs. of clinker 1.08 lbs. of combined water were expelled from the hydrated silica contained in the slurry. Hence column (6) of the annexed Table I. is derived by multiplying the weight of clinker in lbs. by the factor 0.0108. It shows the weight of water expelled from the silica during the production of x lbs. of clinker.

TABLE I

Lbs of Dry Slurry Used and Lbs of Carbon Dioxide and Moisture Liberated therefrom in Producing Different Weights of Clinker Expressed in Lbs

| (1) Lbs of Clinker Produced per 1 Lb of Standard Coal W Lbs | (2) Equivalent Weight of Dry Slurry per 1 Lb of Standard Coal W x 1.5603 Lbs | (3) Lbs of CO ₂ Evolved from the Slurry W x 0.5238 Lb | (4) Lbs of Combined Water Expelled from the Kaolin in the Dry Slurry W x 0.0257 Lb | (5) Lbs of Combined Water Expelled from the Hydrated Silica in the Dry Slurry W x 0.0108 Lb |
|---|--|---|--|---|
| 15 732 | 24 547 | 8 240 | 0.4043 | 0 1699 |
| 15 414 | 24 050 | 8 074 | 0.3961 | 0 1665 |
| 15 102 | 23 564 | 7 910 | 0.3881 | 0 1631 |
| 14 793 | 23 081 | 7 748 | 0.3802 | 0 1598 |
| 14 485 | 22 601 | 7 587 | 0.3723 | 0 1564 |
| 14 177 | 22 120 | 7 426 | 0.3643 | 0 1531 |
| 13 873 | 21 646 | 7 267 | 0.3565 | 0 1498 |
| 13 575 | 21 181 | 7 110 | 0.3489 | 0 1466 |
| 13 271 | 20 707 | 6 951 | 0.3411 | 0 1433 |
| 12 972 | 20 240 | 6 795 | 0.3334 | 0 1401 |
| 12 675 | 19 777 | 6 639 | 0.3257 | 0 1369 |
| 12 386 | 19 326 | 6 488 | 0.3183 | 0 1338 |
| 12 094 | 18 870 | 6 335 | 0.3108 | 0 1306 |
| 11 806 | 18 421 | 6 184 | 0.3034 | 0 1275 |
| 11 522 | 17 978 | 6 035 | 0.2961 | 0 1244 |
| 11 239 | 17 536 | 5 887 | 0.2888 | 0 1214 |
| 10 954 | 17 092 | 5 738 | 0.2815 | 0 1185 |
| 10 675 | 16 656 | 5 592 | 0.2743 | 0 1153 |
| 10 397 | 16 222 | 5 446 | 0.2672 | 0 1123 |
| 10 125 | 15 798 | 5 303 | 0.2602 | 0 1094 |
| 9 8498 | 15 368 | 5 159 | 0.2531 | 0 1064 |
| 9 580 | 14 948 | 5 018 | 0.2462 | 0 1035 |
| 9 311 | 14 528 | 4 877 | 0.2393 | 0 1006 |
| 9 045 | 14 113 | 4 738 | 0.2325 | 0 0977 |
| 8 785 | 13 707 | 4 602 | 0.2258 | 0 0949 |
| 8 680 | 13 543 | 4 547 | 0.2231 | 0 0937 |
| 8 408 | 13 119 | 4 404 | 0.2161 | 0 0908 |
| 7 984 | 12 457 | 4 187 | 0.2052 | 0 0862 |
| 7 570 | 11 811 | 3 965 | 0.1945 | 0 0818 |
| 7 165 | 11 180 | 3 753 | 0.1841 | 0 0774 |
| 6 761 | 10 549 | 3 541 | 0.1738 | 0 0730 |
| 6 359 | 9 922 | 3 331 | 0.1634 | 0 0687 |
| 5 963 | 9 304 | 3 123 | 0.1532 | 0 0644 |
| 5 568 | 8 688 | 2 917 | 0.1431 | 0 0601 |
| 5 181 | 8 084 | 2 714 | 0.1337 | 0 0560 |
| 4 795 | 7 487 | 2 512 | 0.1232 | 0 0518 |
| 4 413 | 6 886 | 2 312 | 0.1134 | 0 0477 |
| 4 032 | 6 291 | 2 117 | 0.1036 | 0 0435 |
| 3 693 | 6 230 | 2 091 | 0.1026 | 0 0431 |
| 3 956 | 6 173 | 2 072 | 0.1017 | 0 0477 |

TABLE I.—(Continued)

| (1) Lbs. of Clinker Produced per 1 Lb. of Standard Coal. W. Lbs. | (2) Equivalent Weight of Dry Slurry per 1 Lb. of Standard Coal. W × 1.5603 Lbs. | (3) Lbs. of CO ₂ Evolved from the Slurry. W × 0.5238 Lb. | (4) Lbs. of Combined Water Expelled from the Kaolin in the Dry Slurry. W × 0.0257 Lb. | (5) Lbs. of Combined Water Expelled from the Hydrated Silica in the Dry Slurry. W × 0.0108 Lb. |
|--|---|--|---|--|
| 3.919 | 6.115 | 2.053 | 0.1007 | 0.0423 |
| 3.882 | 6.057 | 2.033 | 0.0998 | 0.0419 |
| 3.843 | 5.996 | 3.013 | 0.0988 | 0.0415 |
| 3.806 | 5.939 | 1.994 | 0.0978 | 0.0411 |
| 3.769 | 5.881 | 1.974 | 0.0969 | 0.0407 |
| 3.731 | 5.821 | 1.954 | 0.0959 | 0.0403 |
| 3.693 | 5.762 | 1.934 | 0.0949 | 0.0399 |
| 3.656 | 5.704 | 1.915 | 0.0940 | 0.0395 |
| 3.619 | 5.647 | 1.896 | 0.0930 | 0.0391 |
| 3.582 | 5.589 | 1.876 | 0.0921 | 0.0387 |
| 3.545 | 5.531 | 1.857 | 0.0911 | 0.0383 |
| 3.507 | 5.472 | 1.837 | 0.0901 | 0.0379 |
| 3.469 | 5.413 | 1.817 | 0.0892 | 0.0375 |
| 3.432 | 5.355 | 1.798 | 0.0882 | 0.0371 |
| 3.395 | 5.297 | 1.778 | 0.0873 | 0.0367 |
| 3.359 | 5.241 | 1.759 | 0.0863 | 0.0363 |
| 3.320 | 5.180 | 1.739 | 0.0853 | 0.0359 |
| 3.283 | 5.129 | 1.720 | 0.0844 | 0.0355 |
| 2.915 | 4.548 | 1.527 | 0.0749 | 0.0315 |
| 2.549 | 3.977 | 1.335 | 0.0655 | 0.0275 |
| 2.185 | 3.409 | 1.145 | 0.0562 | 0.0236 |
| 1.824 | 2.846 | 0.955 | 0.0469 | 0.0197 |
| 1.466 | 2.287 | 0.768 | 0.0377 | 0.0158 |
| 1.111 | 1.733 | 0.582 | 0.0286 | 0.0120 |
| 0.760 | 1.186 | 0.398 | 0.0195 | 0.0082 |
| 0.411 | 0.641 | 0.215 | 0.0106 | 0.0044 |
| 0.065 | 0.101 | 0.034 | 0.0017 | 0.0007 |
| 0.000 | 0.000 | 0.000 | 0.0000 | 0.0000 |

CHAPTER XIX

CALCULATION OF THE EXIT TEMPERATURES OF THE GASES FROM A PERFECT ROTARY KILN, USING INCOMING AIR PREHEATED TO DIFFERENT TEMPERATURES BY THE OUTGOING CLINKER, NO EXTERNAL OR INTERNAL RADIATION LOSSES OCCURRING

§ 1. ALTHOUGH a perfect rotary kiln is unattainable in practice, it is important to know the conditions which would prevail in a perfect kiln, as in that way we can ascertain how far the kilns actually in use fall short of the ideal, and what are the precise limits to possible improvement in any given particular.

§ 2. We must now define what we mean by a *perfect cement rotary kiln*. In such a kiln:—

- (a) No radiation losses will occur.
- (b) All the heat is communicated to the raw material by conduction and convection of the hot gases produced in the kiln by the combustion of coal.
- (c) The gas leaves the decarbonating zone at $1,481^{\circ}$ F., being at the same temperature as the raw material all the way down the kiln, so that the interchange of heat between the gas and raw material is perfect.
- (d) The slurry is dry, no mechanically mixed water being present.
- (e) The weight of air used per 1 lb. of standard coal (of 12,600 B.Th.U.'s per lb.) is 10.478 lbs., as shown in Chapter X.
- (f) The weight of the combustion gas produced per 1 lb. of standard coal is 11.278 lbs. and the composition is that shown in Chapter X.
- (g) The maximum temperature to which the incoming air can be heated by the outgoing clinker is $2,500^{\circ}$ F. ($1,371^{\circ}$ C.), which is taken as the clinkering temperature.
- (h) The temperature of the external atmosphere is taken as 60° F. (15.6° C.).

§ 3. With these premises we will now proceed to calculate the conditions regulating the temperature of the exit gas when the air is preheated to different temperatures by the outgoing clinker.

This can only be done by calculating a series of heat balances, taking the passage of the combustion gas right down the kiln from the hot end to the cold end. This we will now proceed to do.

§ 4. Let us take the simplest case first, and suppose that the air entering the kiln is not being preheated at all by the outgoing clinker, as shown in Fig. 1.

Hence the 10.478 lbs. of cold air at 60° F. meet with 1 lb. of coal dust, which burns, and as no radiation losses of any kind occur, there are generated 11.278 lbs. of combustion gas at a flame temperature of $3,769^{\circ}$ F., as was explained in Chapters X., XI., and XVI.

This gas then travels through the clinkering and decarbonating zone of the kiln, and escapes from the latter at a temperature of 1481°F , after having given up to the raw material in these zones $7,972\text{ B Th U 's}$

As shown in Chapter XIV, these $7,972\text{ B Th U 's}$ will produce in the clinkering zone $\frac{7972}{918.6} = 8\,680\text{ lbs}$ of clinker

As shown in Chapter XVIII (see also Chapter IX), this weight of clinker is produced from $13\,543\text{ lbs}$ of dry slurry, and there is expelled therefrom $4\,547\text{ lbs}$ of CO_2 at $1,481^{\circ}\text{F}$, also $0\,2231\text{ lbs}$ of H_2O from the kaolin (at $1,472^{\circ}\text{F}$), and $0\,0937\text{ lb}$ of H_2O from the hydrated silica (expelled at 752°F)

These hot gases now pass down the kiln and out of the decarbonating zone into the dehydrating and preheating zone, and finally escape at cold end of the kiln at an exit temperature T , which in this particular case is known by a preliminary calculation to be 212°F . In passing down the kiln these gaseous products yield

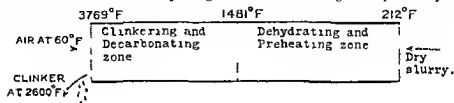


FIG. 1

up their heat to the raw material, and the quantities of heat thus liberated may be calculated as follows —

(a) The heat evolved by $11\,278\text{ lbs}$ of combustion gas sinking from $1,481^{\circ}\text{F}$ to 212°F is

$$11\,278 \times 0.2610 \times (1481 - 212) = 3735.4\text{ B Th U 's}$$

(b) The heat evolved by the $4\,547\text{ lbs}$ of CO_2 evolved from the slurry sinking from $1,481^{\circ}\text{F}$ to 212°F is from our heat tables —

$$4\,547(360.911 - 37.483) = 1470.63\text{ B Th U 's}$$

(c) The heat evolved by the 0.2231 lb of H_2O vapour evolved from the kaolin at $1,472^{\circ}\text{F}$ and sinking to 212°F without condensing to liquid is from our heat tables —

$$0.2231(691.779 - 84.256) = 135.54\text{ B Th U 's}$$

(d) Heat evolved by the 0.0937 lb of H_2O expelled from the hydrated silica at 752°F (400°C) and sinking from that temperature to 212°F without condensing to water will be

$$0.0937(336.257 - 84.256) = 23.61\text{ B Th U 's}$$

(e) Heat evolved by $x\text{ lbs}$ of steam in the furnace gases condensing to liquid water at 212°F , depositing on the cold dry slurry entering the kiln at 60°F , and further cooling from 212° to 60°F in contact with the slurry, is

$$x(970.7 + 212 - 60) = 1122.70x\text{ B Th U 's}$$

Now, obviously —

Heat evolved by furnace gases and steam passing down the kiln = heat absorbed in heating the $13\,543\text{ lbs}$ of slurry from 60° to 1481°F (the temperature at which the latter enters the decarbonating zone)

$$= \text{weight of clinker} \times 652.5 \text{ (see Chapter IX)}$$

$$= 8\,680 \times 652.5 = 5663.7\text{ B Th U 's}$$

Hence we have the following equation —

$$3735.4 + 1470.63 + 135.54 + 23.61 + 1122.70x = 5663.7,$$

$$\text{or} \quad x = 0.266\text{ lb of water}$$

Hence in this case the furnace gases pass away at 212° F., carrying with them the bulk of the steam, but depositing 0.266 lb. of water on the cold incoming dry slurry, which it helps to preheat to 212° F.

All the heat in the outgoing hot clinker is wasted, since none is supposed to be restored by the incoming air. Since the 8.680 lbs. of clinker fall away from the kiln at $2,500^{\circ}$ F. and finally cool to 60° F., and since the mean specific heat of the clinker between 60° and $2,500^{\circ}$ F. is 0.2519, the amount of heat lost is

$$8.680 \times 0.2519 \times (2500 - 60) = 5,335 \text{ B.Th.U.'s.}$$

§ 5. Now let us consider the case when the incoming air is preheated by the outgoing clinker to 100° F., all other circumstances being the same as in § 4.

Then, as shown in Chapter XII., Table III., the flame temperature of the 11.278 lbs. of combustion gas will be $3,794^{\circ}$ F., and the respective quantities of the various components formed in the kiln will be (see Chapter XIV., Table I., and Chapter XVIII., Table I.)

| | |
|---|------------|
| Clinker | 8.785 lbs. |
| Dry slurry | 13.707 " |
| CO ₂ evolved from slurry at $1,481^{\circ}$ F. | 4.602 " |
| H ₂ O expelled from kaolin at $1,472^{\circ}$ F. | 0.2258 lb. |
| H ₂ O expelled from silica at 752° F. | 0.0949 " |

On repeating the preceding calculation we arrive at the following result:—

The furnace gases still pass away from the kiln at 212° F., but more water, amounting to 0.3093 lb., is deposited on the dry entering slurry, leaving 0.4594 lb. of water to pass away as steam.

The 8.785 lbs. of clinker produced in passing away from the kiln and falling in temperature from $2,500^{\circ}$ to 60° F. carry away with it

$$8.785 \times 0.2519 \times (2500 - 60) = 5,400 \text{ B.Th.U.'s.}$$

The 10.478 lbs. of entering air are supposed to be preheated to 100° F. by this hot clinker, and so restore to the furnace 96.5 B.Th.U.'s, which amount to 1.8 per cent. of the heat in the clinker.

§ 6. On repeating a succession of these calculations for air preheated to different amounts we get a series of results from which the table on p. 19.4 is compiled. On studying this table the following facts appear:—

(1) When the entering 10.478 lbs. of air per 1 lb. of coal are preheated by the issuing clinker from anything between 60° and 500° F., the exit temperature keeps steady at 212° F. owing to the fact that the gases are condensing their steam, which separates first as boiling water, and serves to preheat the cold entering dry slurry.

(2) When the clinker preheats the entering air to a higher temperature than 500° F., the temperature of the exit gases steadily decreases (all their water having been condensed) until, where the entering air attains a temperature of $1,100^{\circ}$ F., the exit temperature of the gases falls to 60° F. and the gases issue from the kiln at the temperature of the external atmosphere, having given up all their available heat to the entering raw material.

(3) When the entering air is preheated above $1,100^{\circ}$ F., the heat from the hot gas is absorbed so completely in producing clinker that the 11.278 lbs. of combustion gas produced per 1 lb. of coal do not contain sufficient heat to issue from the kiln even at 60° F. So that in this case the entering dry slurry must be preheated to a temperature ranging from 77° F. (when the entering air is at $1,100^{\circ}$ F.) to a temperature as high as 402° F. (when the entering air is at $2,500^{\circ}$ F.).

In this connection see also Chapters XVI. and XVII., where the case is discussed in detail.

§ 7. These conclusions at first sight seem contrary to common sense, since one would naturally expect that the higher the temperature at which the air

enters the kiln, the higher will be the exit temperature of the emergent gases, whereas the exact contrary is the case, the exit temperature of the gases steadily falling as the temperature of the entering air increases

The reason of this, however, will be apparent to anyone who calculates out a series of heat balances on the lines outlined above. The reason of the phenomenon is, of course, that clinker formation is a high temperature process and not a low temperature process. To make 1 lb of clinker requires (see Chapter IX) 918.6 B Th U's absorbed above 1,481° F, but only 652.5 B Th U's absorbed below 1,481° F.

Hence, if the heat in the entering gases is presented to the raw material in such

TABLE I

Showing Exit Temperature, etc., of Gases from a Perfect Rotary Kiln when 10 478 Lbs of Air are Used per 1 Lb of Standard Coal Burnt, and the Entering Air is Preheated by the Clinker to the Temperature shown in Column (1)

| (1) Tempera- ture of Entering Combustion Air (10 478 Lbs per 1 Lb Coal) | (2) Flame Tempera- ture | (3) Weight of Clinker per 1 Lb of Standard Coal | (4) Tons of Standard Coal Consumed per 100 Tons Clinker | (5) Per Cent of Moisture in Slurry | (6) Tempera- ture of Dry Slurry Entering Kiln | (7) Tempera- ture of Exit Gas | (8) Per Cent of Total Heat in Clinker Restored to Kiln by Entering Air |
|---|----------------------------------|--|--|---|---|---|--|
| ° F | ° F | Lbs | | | ° F | ° F | |
| 60 | 3769 | 8 680 | 11 52 | 0 | 60 | 212 | 0.8 |
| 100 | 3794 | 8 785 | 11 38 | 0 | 60 | 212 | 1.8 |
| 200 | 3855 | 9 045 | 11 06 | 0 | 60 | 212 | 6.1 |
| 300 | 3916 | 9 311 | 10 74 | 0 | 60 | 212 | 10.2 |
| 400 | 3977 | 9 580 | 10 44 | 0 | 60 | 212 | 14.1 |
| 500 | 4038 | 9 850 | 10 15 | 0 | 60 | 212 | 17.8 |
| 600 | 4100 | 10 125 | 9 88 | 0 | 60 | 188 | 21.3 |
| 700 | 4160 | 10 397 | 9 62 | 0 | 60 | 167 | 24.7 |
| 800 | 4222 | 10 675 | 9 37 | 0 | 60 | 129 | 27.9 |
| 900 | 4282 | 10 954 | 9 13 | 0 | 60 | 105 | 31.0 |
| 1000 | 4344 | 11 239 | 8 90 | 0 | 60 | 745 | 34.0 |
| 1100 | 4405 | 11 463 | 8 724 | 0 | 77 | 60 | 36.8 |
| 1200 | 4465 | 11 546 | 8 6613 | 0 | 106 | 60 | 39.6 |
| 1300 | 4525 | 11 815 | 8 4638 | 0 | 134 | 60 | 42.2 |
| 1400 | 4586 | 11 990 | 8 3399 | 0 | 161 | 60 | 44.7 |
| 1500 | 4645 | 12 160 | 8 2238 | 0 | 187 | 60 | 47.1 |
| 1600 | 4706 | 12 332 | 8 1089 | 0 | 212 | 60 | 49.5 |
| 1900 | 4936 | 12 83 | 7 7923 | 0 | 282 | 60 | 56.0 |
| 2100 | 5071 | 13 160 | 7 5986 | 0 | 325 | 60 | 60.0 |
| 2300 | 5224 | 13 48 | 7 4214 | 0 | 365 | 60 | 63.6 |
| 2500 | 5370 | 13 79 | 7 2535 | 0 | 402 | 60 | 67.0 |

a manner that there is a high percentage of available high grade heat, this heat will be readily absorbed with a resulting large output of clinker per 1 lb of coal.

On the other hand, if the available heat in the gases contains a large percentage of low grade heat, this heat is useless for forming clinker and comes down the kiln unabsorbed and appears as heat in the exit gases. In this connection the reader should consult Chapter XV again.

Now preheating the entering air to a high temperature simply means increasing the proportion of high grade to low grade heat in the entering gases and

therefore the more complete absorption of the heat in clinker formation in the upper part of the kiln.

No heat whatever is absorbed in clinker formation unless it is presented at a temperature above 1,481° F. (805° C.), most heat simply passing unabsorbed down the kiln and appearing in the exit gases, and is simply wasted except in so far as it preheats the raw material for the clinkering zone and evaporates superabundant water.

Hence the process, when followed thermodynamically, shows the appalling wastage of heat in the present method of employing gases containing such a large proportion of low-grade heat, as shown in Chapter XV.

For fuel economy it is essential to utilise high-grade heat for high-grade thermal work and low-grade heat for low-grade thermal work, and therefore that the entering gases must be heated as highly as possible.

The present chapter, therefore, forms another proof of the statements set forth in Chapter XV.

§ 8. In the preceding sections the amount of air entering the kiln has been kept steady at the normal amount, viz., 10.478 lbs. per 1 lb. of coal.

It has just been shown that when this is the case, and when the temperature of the entering air attains 1,100° F., the high-grade heat is absorbed so completely in forming clinker that the combustion gas does not contain sufficient heat to preheat the slurry to the temperature of the carbonating zone (1,481° F., or 805° C.), and that therefore the slurry must be separately preheated to various amounts in order to enable it to enter the decarbonating zone at the correct temperature.

As shown in Chapters XVI. and XVII., this state of affairs can be remedied by allowing more than the 10.478 lbs. of air to enter the kiln per 1 lb. of coal, provided that this air is preheated by the clinker so as to restore to the furnace all the heat outgoing in the clinker.

In the table on p. 19.6 the effect is shown of dividing the air stream into two parts after the air has removed all the heat from the clinker. 10.478 lbs. of hot air are allowed to enter the clinkering zone with the coal dust and burn them in the usual way. The balance of hot air is allowed to enter the preheating and dehydrating zone below the carbonating zone, and thus preheat the cold slurry to the necessary degree.

When these arrangements are made and the necessary calculations performed, the following table results.

The weight of air which must be supplied in order to remove all the heat from the clinker is calculated as follows:—

The specific heat of the clinker between 60° and 2,500° F. is 0.2519.

Let W = necessary weight of air.

Then

$$W \times \left\{ \begin{array}{l} \text{Mean specific heat} \\ \text{of air between} \\ T^{\circ} \text{ and } 60^{\circ} \end{array} \right\} \times (T - 60) = \text{weight of clinker} \times 0.2519 \times (2500 - 60).$$

Here T = temperature to which the air is preheated, shown in column (1) of the table on p. 19.6.

The weight of clinker is obtained from column (3), and the mean specific heat of the air between the various temperatures is read off from our specific heat tables.

§ 9. It will be seen from this table that under these conditions the kiln is working under a condition of maximum efficiency and all the heat supplied is consumed in the kiln itself.

So that the exit gases escape at 60° F.—the temperature of the air—and the clinker escapes at 60° F., while the kiln, under the most favourable conditions, can produce 15.732 lbs of clinker per 1 lb. of standard coal burnt, or 100 tons of clinker per 6.3565 tons of standard coal.

TABLE II

All Heat Restored from Clinker to Kilm by the Air Entering Preheated to the Temperature shown in Column (1) by the Weight of Clinker shown in Column (3). The Quantity of Fattening Air is Sufficient to Restore to the Kilm All Heat Outgoing in the Clinker 10,478 Lbs of this Hot Air Enter with the Coal Dust in the Clinkering Zone and the Rest Passes Indirectly into the Preheating Zone

| (1) Temperature of Preheated Air T° | (2) Flame Temperature | (3) Weight of Clinker per 1 Lb. Standard Coal | (4) Tons Standard Coal per 100 Tons Clinker | (5) Temperature of Dry Slurry Entering Kilm | (6) Temperature of Exit Gases | (7) Percentage of Heat in Clinker Restored to Kilm by Entering Air | (8) Total Weight of Air per 1 Lb. of Coal | (9) Air Used with 1 Lb of Coal | (10) Air Used for Preheating Slurry |
|--|-----------------------------|--|--|--|---|---|---|---|---|
| ° F | ° F | Lbs | | ° F | ° F | | Lbs | Lbs | Lbs |
| 2100 | 4405 | 12 522 | 8 6790 | 60 | 60 | 100 | 28 5 | 10 478 | 18 022 |
| 2200 | 4465 | 11 806 | 8 4703 | 60 | 60 | 100 | 34 1 | 10 478 | 23 622 |
| 2300 | 4525 | 12 094 | 8 2686 | 60 | 60 | 100 | 24 8 | 10 478 | 14 322 |
| 2400 | 4586 | 12 386 | 8 0736 | 60 | 60 | 100 | 23 5 | 10 478 | 13 022 |
| 2500 | 4645 | 12 675 | 7 8895 | 60 | 60 | 100 | 22 2 | 10 478 | 11 722 |
| 2600 | 4706 | 12 972 | 7 7089 | 60 | 60 | 100 | 21 2 | 10 478 | 10 722 |
| 2700 | 4805 | 13 271 | 7 5353 | 60 | 60 | 100 | 20 3 | 10 478 | 9 822 |
| 2800 | 4862 | 13 575 | 7 3665 | 60 | 60 | 100 | 19 5 | 10 478 | 9 022 |
| 2900 | 4936 | 13 873 | 7 2082 | 60 | 60 | 100 | 18 7 | 10 478 | 8 222 |
| 3000 | 4995 | 14 177 | 7 0537 | 60 | 60 | 100 | 18 1 | 10 478 | 7 622 |
| 3100 | 5071 | 14 485 | 6 9037 | 60 | 60 | 100 | 17 5 | 10 478 | 7 022 |
| 3200 | 5147 | 14 793 | 6 7599 | 60 | 60 | 100 | 17 0 | 10 478 | 6 522 |
| 3300 | 5224 | 15 102 | 6 6216 | 60 | 60 | 100 | 16 5 | 10 478 | 6 022 |
| 3400 | 5301 | 15 414 | 6 4876 | 60 | 60 | 100 | 16 1 | 10 478 | 5 622 |
| 3500 | 5370 | 15 732 | 6 3565 | 60 | 60 | 100 | 15 6 | 10 478 | 5 122 |

CHAPTER XX

LOSS IN CLINKER OUTPUT DUE TO EXTERNAL RADIATION LOSSES FROM A MODERN ROTARY KILN

§ 1. IN Chapter XIX. a perfect kiln was investigated, in which no losses due to either external or internal radiation occurred.

In the ordinary rotary kiln employed in practice, however, there occur very serious thermal losses due to both external and internal radiation losses. How serious are such losses may be gathered from the fact that it is these radiation losses which are the main factors in reducing the output of the kiln from 6.36 tons of standard coal per 100 tons of clinker to, say, 30 tons per 100 tons of clinker (the average figure now prevailing) and render the wet process a practical one.

In the present chapter it is proposed to confine our attention to the loss of clinker output due to *external radiation losses* from the kiln shell to the atmosphere, reserving for the succeeding chapter the treatment of the internal radiation losses.

§ 2. The British Portland Cement Research Association has during the last few years carried out twenty-nine tests of cement rotary kilns, and has thereby acquired an accurate experimental knowledge of the external radiation losses suffered by modern rotary kilns.

The figures vary somewhat widely. But in one test of the Works No. 16, made in 1920, the total radiation loss came out as equivalent to 5.60 tons of standard coal per 100 tons of coal burnt, or 705.6 B.Th.U.'s per 1 lb. of standard coal burnt (of 12,600 B.Th.U.'s per lb.).

In the Works No. 26, carried out in 1924, their corresponding figures were 5.61 tons of standard coal per 100 tons of clinker burnt, or 706.7 B.Th.U.'s per 1 lb. of standard coal burnt (12,600 B.Th.U.'s).

These figures may be assumed as of the order of the external radiation losses which occur in a good modern rotary cement kiln.

As it is necessary to take some definite figures in our calculations, *we will assume that for every 1 lb. of standard coal burnt in the furnace some 700 B.Th.U.'s are wasted in external radiation.*

In other words, for every 12,600 B.Th.U.'s liberated within the kiln, 700 B.Th.U.'s are wasted in heating the air around the kiln.

This amounts to an *apparent loss* of heat only about $5\frac{1}{2}$ per cent.

The real effective loss, however, is nearer 15 per cent., as we will see below, and is, therefore, very serious.

§ 3. In Chapters IX., XIV., and XV. it has been demonstrated that it is the high-grade heat only—*i.e.*, the heat which is available at a temperature *above* $1,481^{\circ}$ F. (805° C.)—which is capable of generating clinker. Low-grade heat—heat available below $1,481^{\circ}$ F.—cannot produce a single ounce of clinker, even if available to the extent of millions of B.Th.U.'s.

Consequently, it is the *loss of high grade heat which is responsible for loss of clinker output*

The loss of low grade heat—no matter how large—does not reduce the clinker output by a single particle, provided always there is sufficient low grade heat available to do the necessary preliminary thermal work of preheating the entering slurry up to the temperature of $1,481^{\circ}\text{F}$ (805°C) preparatory to its entering the CO_2 expulsion and clinkering zone

If there is not sufficient low grade heat present to do this, its place must be taken by high grade heat, to the great detriment of clinker output

This point was developed in detail in Chapter XV. In Chapter IV, it was shown that it requires 918.6 B Th U's of high grade heat (*i.e.*, B Th U's available *above* $1,481^{\circ}\text{F}$ (805°C)) to generate 1 lb. of clinker)

A million B Th U's of low grade heat will not generate a single ounce of clinker

§ 4 Consequently, on examining the thermal losses by external radiation, it is not sufficient to examine them solely by the B Th U's lost, irrespective of the nature of the B Th U's. It makes all the difference in the world to clinker output whether these lost B Th U's come under the category of high grade heat or low-grade heat

Therefore, in order to arrive at the loss of clinker output by external radiation, it is very necessary to examine the *source* of the B Th U's. If the B Th U's are *lost* from the clinkering and decarbonating zones, it is obvious that they represent lost *high grade heat* (as all the gases in these zones are above $1,481^{\circ}\text{F}$), whereas any heat lost in the dehydrating and preheating zones—the colder zones of the clinker—merely represent *low-grade heat* (since the gases and material inside are *below* $1,481^{\circ}\text{F}$), and so these losses are not serious from the clinker output point of view (although for steam generation or evaporative purposes these losses may be considered serious where waste-heat boilers are in use)

Every 918.6 B Th U's lost of high grade heat represent the loss of 1 lb. of potential clinker as shown in Chapters XIV and XV

§ 5 A careful estimation of the radiation losses from different sections of all the cement rotary kilns tested by the Research Association gave the following average figures —

| | Per Cent. |
|--|-----------|
| (1) Radiation loss from clinkering zone | 39 |
| (2) " " CO_2 expulsion zone | 37 |
| (3) " " dehydrating and preheating zones | 21 |
| (4) " " end sections of kilns | 12 |
| | <hr/> 100 |

Hence, if we accept as typical of a modern rotary cement kiln the estimate of the radiation losses made in § 2, *viz.*, 700 B Th U's per 12,600 B Th U's liberated by 1 lb. of standard coal consumed within the kiln, then we arrive at the following figures

Per 1 lb. of standard coal burnt in kiln, and thus liberating 12,600 B Th U's in the kiln, we lose by external radiation —

| | B Th.U's. | |
|---|-------------------------|------------------------|
| (1) In the clinkering zone | $0.39 \times 700 = 273$ | } 46 { High grade heat |
| (2) In the CO_2 expulsion zone | $0.37 \times 700 = 259$ | |
| (3) In the dehydrating and preheating zones | $0.21 \times 700 = 147$ | } 231 { Low grade heat |
| (4) End sections of kiln | $0.12 \times 700 = 84$ | |
| | <hr/> 700 | |

The heat radiated away from both the hot end section of the kiln and the colder end is classified as low-grade heat because in both cases the gases immediately in contact with the ends are certainly at a temperature below $1,481^{\circ}\text{F.}$ (805°C.).

§ 6. It will be seen from § 5 that for every 12,600 B.Th.U.'s liberated within the kiln no less than 469 B.Th.U.'s are lost by external radiation from the clinkering and decarbonating zones, and that the whole of these 469 B.Th.U.'s represent a loss of *high-grade heat*.

Now the loss from these zones of 918.6 B.Th.U.'s of high-grade heat corresponds to a loss of 1 lb. of clinker. So that a loss of 469 B.Th.U.'s corresponds to a loss of $\frac{469}{918.6} = 0.5106$ lb. of clinker.

In other words, per 1 lb. standard coal (of 12,600 B.Th.U.'s per lb.) burnt inside the kiln, we lose the production of 0.510 lb. of clinker through external radiation losses. This is a very serious loss. Thus, take the case of a rotary cement kiln producing 100 lbs. of clinker for every 30.120 lbs. standard coal burnt.

TABLE I

Showing Saving that would be Effected by Stopping All External Radiation from Clinkering and Decarbonating Zone of Kiln Shell

| With Present External Radiation Loss from Clinkering and Decarbonating Zone, of 469 B.Th.U.'s per 1 Lb. of Standard Coal (12,600 B.Th.U.'s per Lb.) Burnt. | | With All External Radiation Losses Stopped from Clinkering and Decarbonating Zone. | |
|--|--|--|---|
| Tons of Standard Coal Burnt for 100 Tons of Clinker Produced. x Tons. | Loss of Clinker Output Due to External Radiation from Clinkering and Decarbonating Zone. $0.5106x$ Tons. | Increased Clinker Output per x Tons of Standard Coal Burnt Due to Stopping All Radiation from Clinkering and Decarbonating Zone. $100 + 0.5106x$. | Tons of Standard Coal Consumed per 100 Tons of Clinker Produced when All External Radiation is Stopped from Clinkering and Decarbonating Zone. $\frac{100x}{100 + 0.5106x}$. |
| 20 | 10.2 | 110.2 | 18.14 |
| 21 | 10.72 | 110.72 | 18.97 |
| 22 | 11.23 | 111.23 | 19.78 |
| 23 | 11.74 | 111.74 | 20.51 |
| 24 | 12.25 | 112.25 | 21.38 |
| 25 | 12.76 | 112.76 | 22.17 |
| 26 | 13.27 | 113.27 | 22.95 |
| 27 | 13.79 | 113.79 | 23.72 |
| 28 | 14.30 | 114.30 | 24.50 |
| 29 | 14.81 | 114.81 | 25.26 |
| 30 | 15.32 | 115.32 | 26.02 |
| 31 | 15.83 | 115.83 | 26.77 |
| 32 | 16.34 | 116.34 | 27.51 |
| 33 | 16.85 | 116.85 | 28.24 |
| 34 | 17.36 | 117.36 | 28.97 |
| 35 | 17.87 | 117.87 | 29.70 |

Then the loss of clinker output by external radiation is

$$30.12 \times 0.5106 = 15.37 \text{ lbs of clinker}$$

In other words, if we stopped all radiation losses in the upper end of the kiln then instead of producing 100 lbs of clinker, we would obtain 115.37 lbs of clinker per 30.12 lbs of coal burnt

So that the 100 lbs of clinker would now be produced by

$$\frac{30.12 \times 100}{115.37} = 26.1 \text{ lbs of coal}$$

The table on p 203 has been calculated to show the serious practical losses which occur due to kiln external radiation

Summary—*In the present rotary kiln each 1 lb of standard coal burnt represents a loss of nearly $\frac{1}{2}$ lb of clinker due to external radiation from the clinkering and CO_2 expulsion zone*

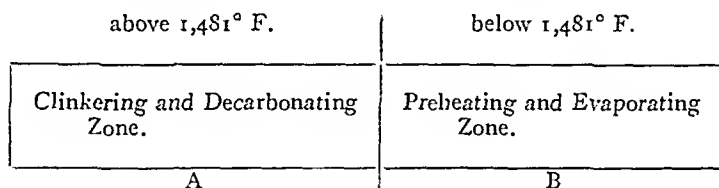
In a cement kiln consuming 30.12 tons of standard coal per 100 tons of clinker produced a stoppage of all external radiation loss from the clinkering and de carbonating zone would reduce the fuel consumption to 26.1 tons of coal per 100 tons of clinker

CHAPTER XXI

LOSS IN CLINKER OUTPUT DUE TO INTERNAL RADIATION LOSSES FROM A MODERN ROTARY KILN

§ 1. THE loss of clinker output from radiation inside the kiln is very much more serious than the loss due to external radiation. This is a source of loss which has been entirely neglected to date.

§ 2. Let us illustrate what we mean by internal radiation.



The rotary kiln may be considered as a closed box divided into two parts, as shown in the figure; the part A consists of the clinkering and CO_2 expulsion zone, while the part B consists of the preheating and evaporating zone. All the heat contained in A is of the high-grade variety, *i.e.*, is available *above* $1,481^{\circ}$ F. (805° C.) (because the gases and solid materials contained in A, as well as the walls of A, are all at a temperature above $1,481^{\circ}$ F.), while the heat contained in B is of the low-grade variety (since all the gases, solid materials contained in B, as well as the walls of B, are at a temperature lower than $1,481^{\circ}$ F.).

Now the part A is at a white heat and filling with very hot gas impregnated with fine dust, while the part B is comparatively cold.

Consequently, there is going on continuously inside the kiln a steady radiation of heat from A into B—from a region of high temperatures to a region of lower temperatures.

But every B.Th.U. thus transmitted across the dividing line between A and B represents a loss of high-grade heat, and every 918.6 B.Th.U.'s of high-grade heat thus lost represent the loss of 1 lb. of clinker, which otherwise could be formed in A, as has been explained in Chapters IX., XIV., XV., and XVII.

So that this internal radiation loss represents a very serious diminution in the capacity of the clinkering and decarbonating zone to produce clinker, and as we will show immediately, the loss of clinker output from this course is immensely greater than the loss from the external radiation. *Indeed, this internal radiation is the main cause of the inefficiency of the modern rotary kiln.*

§ 3. It is now necessary to determine the total amount of the loss. This can be done as follows:—

Consider the clinkering and decarbonating zone of a kiln. In it 1 lb. of

standard coal in burning combines with 10 478 lbs of air, and produces 11 278 lbs of combustion gas, by means of which the heating in the kiln takes place

Now, if the 10 478 lbs of air were at the temperature of the atmosphere and no losses at all occurred, it has been shown in Chapter XII that a flame temperature of 3,769° F would be attained, the 11 278 lbs of combustion gas, in passing down the clinkering and decarbonating zone and giving up its heat to the raw material, finally escapes from the decarbonating zone at 1 481° F (805° C)

This gas, in sinking from 3 769° to 1,481° F, can only yield 7,972 B Th U's above 1,481° F (as shown in Chapter XIV) Hence, under these conditions of no thermal loss of any kind, 1 lb of standard coal can only produce 7,972 B Th U's of high grade heat above 1,481° F and so this is the only amount which is available for clinker formation

But if the 10 478 lbs of air presented to the 1 lb of coal be preheated to any given temperature t° F before it is allowed to combine with the coal, and the amount of heat thus imparted to the 10 478 lbs is Q , B Th U's and this must be added on to the 7 972 B Th U's, which thus became available for clinker formation

This was explained in Chapters XII and XIV, which the reader is recommended to peruse once again

Hence the net result is this —

If no thermal loss of any kind occurs in the clinkering and decarbonating zone, there is available for 1 lb of coal for clinker formation 7,972 B Th U's of high grade heat due to the heat from the coal together with Q , B Th U's due to the heat already contained in the 10 478 lbs of hot air which is preheated to a temperature t° above that of the atmosphere (60° F)

I.e., if Q_{air} be the total available high grade heat in the clinkering and decarbonating zone, and it be then contained in 10 478 lbs of entering air—

$$10\,478 \text{ lbs of entering air } Q_{\text{air}} = 7972 + Q \quad (1)$$

§ 4 But this total amount of high grade heat developed in the clinkering and decarbonating zone must supply the following losses —

- (a) The heat required to expel the CO_2 from the carbonate at 1 481° F (805° C), and form the clinker and raise the mass to the clinkering temperature of 2,500° F In Chapter IX this amount of heat was shown to equal 918.6 B Th U's per 1 lb of clinker formed
- (b) The heat radiated externally from the clinkering and decarbonating zone per 1 lb of standard coal burnt, let us call this R This has been shown in Chapter XX to amount to 469 B Th U's per 1 lb of standard coal burnt (of 12,600 B Th U's per lb)
- (c) The heat radiated from the clinkering and decarbonating zone to the lower part of the zone This has to be determined Let it be R_1 B Th U's per 1 lb of standard coal burnt in the kiln

Then we have the following equation to determine this internal radiation —

DIAGRAM

External radiation loss from clinkering and decarbonating zone per 1 lb. of standard coal burnt = R_E .

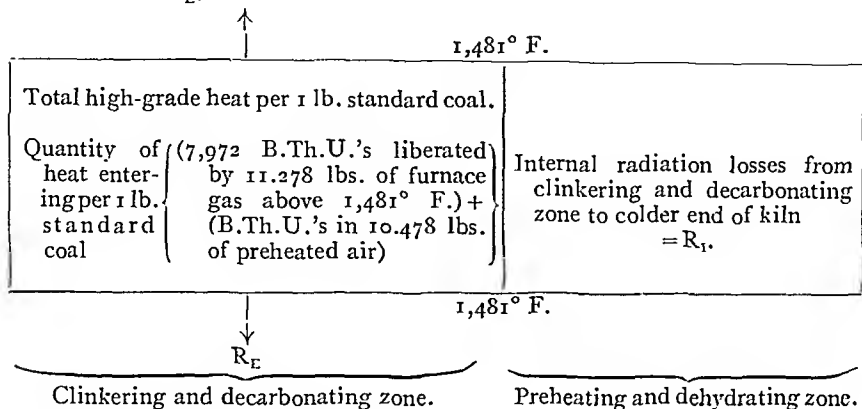


Diagram illustrating the equation:—

| | | | | | | |
|-----------------------------|---|-------------------------------|---|-------------------------------|---|--|
| Total High-grade Heat | = | Internal Radiation Loss | + | External Radiation Loss | + | Heat Absorbed in Clinker Formation |
| $7972 + Q_i$ | = | R_i | + | R_E | + | 918.6W |

$7972 + \left\{ \begin{array}{l} \text{B.Th.U.'s in preheated} \\ \text{10.478 lbs. of air} \end{array} \right\} = \text{B.Th.U.'s absorbed in forming the clinker}$
 + External radiation from clinkering and decarbonating zone per 1 lb. of coal burnt
 + Internal radiation from clinkering and decarbonating zone per 1 lb. of coal burnt;

or, in symbols, if W = lbs. of clinker formed per 1 lb. of coal—

$$7972 + Q_i = W \times 918.6 + R_E + R_i \quad (1)$$

whence

$$R_i + R_E = 7972 + Q_i - W \times 918.6 \quad (2)$$

All these quantities are known in a modern kiln, so that it is possible to calculate the loss of internal radiation R_i for a given clinker output.

§ 5. As the external radiation from the clinkering and decarbonating zone varies for different kilns, we will first calculate the total radiation loss with any given output, their total radiation losses embracing both the internal and external losses.

From the preceding relationship of § 4 we find:—

Internal radiation + External radiation per 1 lb. of standard coal burnt

$$= (7972 + \text{B.Th.U.'s in preheated air}) = \left\{ \begin{array}{l} \text{Weight of clinker} \\ \text{formed per 1 lb.} \\ \text{of coal} \end{array} \right\} \times 918.6;$$

or

$$R_E + R_i = (7972 + Q_i) - W \times 918.6 \quad (3)$$

§ 6. The value of Q_i (the heat contained in the 10.478 lbs. of preheated air necessary for 1 lb. of coal burnt) is calculated for different temperatures in Chapter XII., Table I.

The values of the heat absorbed in forming different weights of clinker ($W \times 918.6$ B.Th.U.'s per W lbs. clinker formed) are calculated in Chapter XIV., Table I., column (2), the corresponding amount of clinker producible being given in column (3) of the same table.

Showing Combined Internal and External Radiation Losses from the Clinkering the External Atmosphere, expressed in B Th U's per 1 lb of

Calculated from the formula —

$$R_t + R_e = 7972 + Q_r - W \times 918.6$$

| (1) Clinker Production in Kilo | | (2) Temperature ° of Entering Air ° F | (3) 60° F | (4) 100° F | (5) 200° F | (6) 300° F |
|--|--|--|--------------|----------------|---------------|---------------|
| Tons Standard Coal Consumed per 100 Tons Clinker Produced | Lbs Clinker Produced per 1 Lb Standard Coal Burnt | Quantity of Heat in 10 478 Lbs. of Preheated Air } Q_r | B Th U ° | B Th U 96.5 | B Th U 338 | B Th U 582 |
| 6 356 | 15 732 | Total Radiation $R_t + R_e$ | 0 | 0 | 0 | 0 |
| 6 488 | 15 414 | Total Radiation $R_t + R_e$ | 0 | 0 | 0 | 0 |
| 6 622 | 15 102 | Total Radiation $R_t + R_e$ | 0 | 0 | 0 | 0 |
| 6 760 | 14 793 | Total Radiation $R_t + R_e$ | 0 | 0 | 0 | 0 |
| 6 904 | 14 485 | Total Radiation $R_t + R_e$ | 0 | 0 | 0 | 0 |
| 7 054 | 14 177 | Total Radiation $R_t + R_e$ | 0 | 0 | 0 | 0 |
| 7 208 | 13 873 | Total Radiation $R_t + R_e$ | 0 | 0 | 0 | 0 |
| 7 367 | 13 575 | Total Radiation $R_t + R_e$ | 0 | 0 | 0 | 0 |
| 7 535 | 13 271 | Total Radiation $R_t + R_e$ | 0 | 0 | 0 | 0 |
| 7 709 | 12 972 | Total Radiation $R_t + R_e$ | 0 | 0 | 0 | 0 |
| 7 889 | 12 675 | Total Radiation $R_t + R_e$ | 0 | 0 | 0 | 0 |
| 8 074 | 12 386 | Total Radiation $R_t + R_e$ | 0 | 0 | 0 | 0 |
| 8 269 | 12 094 | Total Radiation $R_t + R_e$ | 0 | 0 | 0 | 0 |
| 8 470 | 11 806 | Total Radiation $R_t + R_e$ | 0 | 0 | 0 | 0 |

and Decarbonating Zone into the Preheating and Dehydrating Zone, and into Standard Coal (of 12,600 B.Th.U.'s) burnt in the Cement Rotary Kiln

Where $\left\{ \begin{array}{l} R_1 = \text{internal radiation loss from clinkering and decarbonating zone into preheating and dehydrating zone, expressed in B.Th.U.'s per 1 lb. of standard coal consumed in the furnace.} \\ R_E = \text{external radiation loss from the clinkering and decarbonating zone into the air expressed in B.Th.U.'s per 1 lb. of coal burnt.} \\ Q_c = \text{B.Th.U.'s brought into furnace by 10.478 lbs. of the preheated air.} \\ W = \text{lbs. of clinker produced per 1 lb. of standard coal burnt in the furnace.} \end{array} \right.$

| (7) | (8) | (9) | (10) | (11) | (12) | (13) | (14) | (15) | (16) |
|----------------|-----------------|-----------------|-----------------|-----------------|-----------------|-----------------|-----------------|-----------------|-----------------|
| 400° F. | 500° F. | 600° F. | 700° F. | 800° F. | 900° F. | 1000° F. | 1500° F. | 2000° F. | 2500° F. |
| B.Th.U. 828 | B.Th.U. 1076 | B.Th.U. 1327 | B.Th.U. 1579 | B.Th.U. 1834 | B.Th.U. 2090 | B.Th.U. 2349 | B.Th.U. 3673 | B.Th.U. 5050 | B.Th.U. 6479 |
| 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 291 |
| 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 578 |
| 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 862 |
| 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1145 |
| 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1428 |
| 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 278 | 1707 |
| 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 552 | 1981 |
| 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 831 | 2260 |
| 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1107 | 2536 |
| 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1376 | 2808 |
| 0 | 0 | 0 | 0 | 0 | 0 | 0 | 267 | 1644 | 3073 |
| 0 | 0 | 0 | 0 | 0 | 0 | 0 | 535 | 1912 | 3341 |
| 0 | 0 | 0 | 0 | 0 | 0 | 0 | 800 | 2177 | 3606 |

| (1) Clinker Production in kiln | | (2) Temperature* of Entering Air ° F | (3) 60° F | (4) 100° F | (5) 200° F | (6) 300° F |
|--|---|--|--------------|----------------|---------------|---------------|
| Tons Standard Coal Consumed per 100 Tons Clinker Produced | Lbs Clinker Produced per 1 Lb Standard Coal Burned | Quantity of Heat in 20 478 Lbs. of Preheated Air } Q_r | B Th U ° | B Th U 96.5 | B Th U 338 | B Th U 582 |
| 8 679 | 11 522 | Total Radiation $R_R + R_I$ | 0 | 0 | 0 | 0 |
| 8 898 | 11 239 | Total Radiation $R_R + R_I$ | 0 | 0 | 0 | 0 |
| 9 129 | 10 954 | Total Radiation $R_R + R_I$ | 0 | 0 | 0 | 0 |
| 9 368 | 10 675 | Total Radiation $R_R + R_I$ | 0 | 0 | 0 | 0 |
| 9 618 | 10 397 | Total Radiation $R_R + R_I$ | 0 | 0 | 0 | 0 |
| 9 876 | 10 125 | Total Radiation $R_R + R_I$ | 0 | 0 | 0 | 0 |
| 10 153 | 9 850 | Total Radiation $R_R + R_I$ | 0 | 0 | 0 | 0 |
| 10 438 | 9 580 | Total Radiation $R_R + R_I$ | 0 | 0 | 0 | 0 |
| 10 740 | 9 311 | Total Radiation $R_R + R_I$ | 0 | 0 | 0 | 0 |
| 11 056 | 9 045 | Total Radiation $R_R + R_I$ | 0 | 0 | 0 | 245 |
| 11 383 | 8 785 | Total Radiation $R_R + R_I$ | 0 | 0 | 240 | 484 |
| 11 521 | 8 680 | Total Radiation $R_R + R_I$ | 0 | 96.5 | 338 | 582 |
| 11 893 | 8 408 | Total Radiation $R_R + R_I$ | 254 | 350.5 | 592 | 836 |
| 12 325 | 7 984 | Total Radiation $R_R + R_I$ | 638 | 734.5 | 976 | 1220 |
| 13 210 | 7 570 | Total Radiation $R_R + R_I$ | 1018 | 1114.5 | 1356 | 1600 |
| 13 957 | 7 165 | Total Radiation $R_R + R_I$ | 1390 | 1486.5 | 1728 | 1972 |
| 14 791 | 6 761 | Total Radiation $R_R + R_I$ | 1762 | 1858.5 | 2100 | 2344 |
| 15 726 | 6 359 | Total Radiation $R_R + R_I$ | 2131 | 2227.5 | 2469 | 2713 |
| 16 770 | 5 963 | Total Radiation $R_R + R_I$ | 2494 | 2590.5 | 2832 | 3076 |

| (1) Clinker Production in Lbs. | | (2) Temperature* of Entering Air ° F | (3) 60° F | (4) 100° F | (5) 200° F | (6) 300° F |
|--|--|--|--------------|----------------|---------------|---------------|
| Tons Standard Coal Consumed per 100 Tons Clinker Produced | Lbs. Clinker Produced per 1 Lb Standard Coal Burned | Quantity of Heat in 10 478 Lbs. of Preheated Air } Q_r | B Th U ° | B Th U 96 S | B Th U 338 | B Th U 582 |
| 17 960 | 5 568 | Total Radiation $R_R + R_I$ | 2857 | 2953 5 | 3195 | 3439 |
| 19 301 | 5 181 | Total Radiation $R_R + R_I$ | 3213 | 3309 5 | 3551 | 3795 |
| 20 855 | 4 795 | Total Radiation $R_R + R_I$ | 3567 | 3663 5 | 3905 | 4149 |
| 22 660 | 4 413 | Total Radiation $R_R + R_I$ | 3919 | 4015 5 | 4257 | 4501 |
| 24 801 | 4 030 | Total Radiation $R_R + R_I$ | 4268 | 4364 5 | 4606 | 4850 |
| 25 044 | 3 993 | Total Radiation $R_R + R_I$ | 4304 | 4400 5 | 4642 | 4886 |
| 25 278 | 3 956 | Total Radiation $R_R + R_I$ | 4338 | 4434 5 | 4676 | 4920 |
| 25 517 | 3 919 | Total Radiation $R_R + R_I$ | 4372 | 4468 5 | 4710 | 4954 |
| 25 760 | 3 882 | Total Radiation $R_R + R_I$ | 4406 | 4502 5 | 4744 | 4988 |
| 26 021 | 3 843 | Total Radiation $R_R + R_I$ | 4442 | 4538 5 | 4780 | 5024 |
| 26 274 | 3 806 | Total Radiation $R_R + R_I$ | 4476 | 4572 5 | 4814 | 5058 |
| 26 532 | 3 760 | Total Radiation $R_R + R_I$ | 4510 | 4606 5 | 4848 | 5092 |
| 26 802 | 3 731 | Total Radiation $R_R + R_I$ | 4545 | 4641 5 | 4883 | 5127 |
| 27 078 | 3 693 | Total Radiation $R_R + R_I$ | 4580 | 4676 5 | 4918 | 5162 |
| 27 352 | 3 656 | Total Radiation $R_R + R_I$ | 4614 | 4710 5 | 4952 | 5196 |
| 27 632 | 3 619 | Total Radiation $R_R + R_I$ | 4648 | 4744 5 | 4986 | 5230 |
| 27 917 | 3 582 | Total Radiation $R_R + R_I$ | 4682 | 4778 5 | 5020 | 5264 |
| 28 209 | 3 545 | Total Radiation $R_R + R_I$ | 4716 | 4812 5 | 5054 | 5298 |
| 28 514 | 3 507 | Total Radiation $R_R + R_I$ | 4751 | 4847 5 | 5089 | 5333 |

I.—(Continued)

| (7) 400° F | (8) 500° F | (9) 600° F | (10) 700° F | (11) 800° F | (12) 900° F | (13) 1000° F | (14) 1500° F | (15) 2000° F | (16) 2500° F |
|---------------|----------------|----------------|----------------|----------------|----------------|-----------------|-----------------|-----------------|-----------------|
| B Th U 828 | B Th U 1076 | B Th U 1327 | B Th U 1579 | B Th U 1834 | B Th U 2090 | B Th U 2349 | B Th U 3673 | B Th U 5050 | B Th U 6479 |
| 3685 | 3933 | 4184 | 4436 | 4691 | 4947 | 5206 | 6530 | 7907 | 9,336 |
| 4041 | 4289 | 4540 | 4792 | 5047 | 5303 | 5562 | 6886 | 8263 | 9,692 |
| 4395 | 4643 | 4894 | 5146 | 5401 | 5657 | 5916 | 7240 | 8617 | 10,046 |
| 4747 | 4995 | 5246 | 5498 | 5753 | 6009 | 6268 | 7592 | 8969 | 10,398 |
| 5096 | 5344 | 5595 | 5847 | 6102 | 6358 | 6617 | 7941 | 9318 | 10,747 |
| 5132 | 5380 | 5631 | 5883 | 6138 | 6394 | 6653 | 7977 | 9354 | 10,783 |
| 5166 | 5414 | 5665 | 5917 | 6172 | 6428 | 6687 | 8011 | 8388 | 10,817 |
| 5200 | 5448 | 5699 | 5951 | 6206 | 6462 | 6721 | 8045 | 9422 | 10,851 |
| 5234 | 5482 | 5733 | 5985 | 6240 | 6496 | 6755 | 8079 | 9456 | 10,885 |
| 5270 | 5518 | 5769 | 6021 | 6276 | 6532 | 6791 | 8115 | 9492 | 10,921 |
| 5304 | 5552 | 5803 | 6055 | 6310 | 6566 | 6825 | 8149 | 9526 | 10,955 |
| 5330 | 5586 | 5837 | 6089 | 6344 | 6600 | 6859 | 8183 | 9560 | 10,989 |
| 5373 | 5621 | 5872 | 6124 | 6379 | 6635 | 6894 | 8218 | 9595 | 11,024 |
| 5408 | 5656 | 5907 | 6159 | 6414 | 6670 | 6929 | 8253 | 9630 | 11,059 |
| 5442 | 5690 | 5941 | 6193 | 6448 | 6704 | 6963 | 8287 | 9664 | 11,093 |
| 5476 | 5724 | 5975 | 6227 | 6482 | 6738 | 6997 | 8321 | 9698 | 11,127 |
| 5510 | 5758 | 6009 | 6261 | 6516 | 6772 | 7031 | 8355 | 9732 | 11,161 |
| 5544 | 5792 | 6043 | 6295 | 6550 | 6806 | 7065 | 8389 | 9766 | 11,195 |
| 5571 | 5827 | 6078 | 6330 | 6585 | 6841 | 7100 | 8424 | 9801 | 11,230 |

TABLE

| (1) Clinker Production in Lbs | | (2) Temperature ° of Enter 25° Air $^{\circ}$ F | (3) 60° F | (4) 500° F | (5) 200° F | (6) 300° F |
|--|--|---|-----------------------|------------------------|------------------------|------------------------|
| Tons Standard Coal Consumed per 100 Tons Clinker Produced | Lbs Clinker Produced per 1 Lb Standard Coal Burnt | Quantity of Heat to 20 478 Lbs of Preheated Air } Q_r | B Th U ° | B Th U 96.5 | B Th U 33.8 | B Th U 58.2 |
| 28 827 | 3 469 | Total Radiation $R_g + R_i$ | 4785 | 4881.5 | 5123 | 5367 |
| 29 137 | 3 432 | Total Radiation $R_g + R_i$ | 4819 | 4915.5 | 5157 | 5401 |
| 29 455 | 3 395 | Total Radiation $R_g + R_i$ | 4853 | 4949.5 | 5191 | 5435 |
| 29 771 | 3 359 | Total Radiation $R_g + R_i$ | 4887 | 4983.5 | 5225 | 5469 |
| 30 120 | 3 320 | Total Radiation $R_g + R_i$ | 4922 | 5018.5 | 5260 | 5504 |
| 30 460 | 3 283 | Total Radiation $R_g + R_i$ | 4956 | 5052.5 | 5294 | 5538 |
| 34 305 | 2 915 | Total Radiation $R_g + R_i$ | 5295 | 5391.5 | 5633 | 5877 |
| 39 246 | 2 549 | Total Radiation $R_g + R_i$ | 5631 | 5727.5 | 5969 | 6213 |
| 45 766 | 2 185 | Total Radiation $R_g + R_i$ | 5965 | 6061.5 | 6303 | 6547 |
| 54 824 | 1 824 | Total Radiation $R_g + R_i$ | 6296 | 6392.5 | 6634 | 6878 |
| 70 126 | 1 466 | Total Radiation $R_g + R_i$ | 6625 | 6721.5 | 6963 | 7207 |
| 90 000 | 1 111 | Total Radiation $R_g + R_i$ | 6951 | 7047.5 | 7289 | 7533 |
| 131 579 | 0 760 | Total Radiation $R_g + R_i$ | 7274 | 7370.5 | 7612 | 7856 |
| 243 393 | 0 411 | Total Radiation $R_g + R_i$ | 7594 | 7690.5 | 7932 | 8176 |
| 1538 461 | 0 065 | Total Radiation $R_g + R_i$ | 7912 | 8008.5 | 8250 | 8494 |
| | 0 000 | Total Radiation $R_g + R_i$ | 7972 | 8068.5 | 8310 | 8554 |

I — (Continued)

| (7) 400° F | (8) 500° F | (9) 600° F | (10) 700° F | (11) 800° F | (12) 900° F | (13) 1000° F | (14) 1500° F | (15) 2000° F | (16) 2500° F |
|---------------|----------------|----------------|----------------|----------------|----------------|-----------------|-----------------|-----------------|-----------------|
| B Th U 828 | B Th U 1076 | B Th U 1327 | B Th U 1579 | B Th U 1834 | B Th U 2090 | B Th U 2349 | B Th U 3673 | B Th U 5050 | B Th U 6479 |
| 5613 | 5861 | 6112 | 6364 | 6619 | 6,875 | 7,134 | 8,458 | 9,835 | 11,264 |
| 5647 | 5895 | 6146 | 6398 | 6653 | 6,909 | 7,168 | 8,492 | 9,869 | 11,298 |
| 5681 | 5929 | 6180 | 6432 | 6687 | 6,943 | 7,202 | 8,526 | 9,903 | 11,332 |
| 5715 | 5963 | 6214 | 6466 | 6721 | 6,977 | 7,236 | 8,560 | 9,937 | 11,376 |
| 5750 | 5998 | 6249 | 6501 | 6756 | 7,012 | 7,271 | 8,595 | 9,972 | 11,401 |
| 5784 | 6032 | 6283 | 6535 | 6790 | 7,046 | 7,305 | 8,629 | 10,006 | 11,435 |
| 6123 | 6371 | 6622 | 6874 | 7129 | 7,385 | 7,644 | 8,968 | 10,345 | 11,774 |
| 6459 | 6707 | 6958 | 7210 | 7465 | 7,721 | 7,980 | 9,304 | 10,681 | 12,110 |
| 6793 | 7041 | 7292 | 7544 | 7799 | 8,055 | 8,314 | 9,638 | 11,015 | 12,444 |
| 7124 | 7372 | 7623 | 7875 | 8130 | 8,386 | 8,645 | 9,969 | 11,346 | 12,775 |
| 7453 | 7701 | 7952 | 8204 | 8459 | 8,715 | 8,974 | 10,298 | 11,675 | 13,104 |
| 7779 | 8027 | 8278 | 8530 | 8785 | 9,041 | 9,300 | 10,624 | 12,001 | 13,430 |
| 8102 | 8350 | 8601 | 8853 | 9108 | 9,364 | 9,623 | 10,947 | 12,324 | 13,753 |
| 8422 | 8670 | 8921 | 9173 | 9428 | 9,684 | 9,943 | 11,267 | 12,644 | 14,073 |
| 8740 | 8988 | 9239 | 9491 | 9746 | 10,002 | 10,261 | 11,585 | 12,962 | 14,391 |
| 8800 | 9048 | 9299 | 9551 | 9806 | 10,062 | 10 321 | 11,645 | 13,022 | 14,451 |

In a perfect kiln this quantity of heat is obviously identical with the number of B Th U 's liberated by 11 778 lbs of combustion gas falling from the maximum flame temperatures produced where the air necessary for combustion is preheated to different temperatures $t^{\circ}\text{C}$

§ 7 From the values of QW and 918 6W contained in the above-mentioned table, the preceding Table I is calculated

It gives the values of the sum of the losses of *high grade heat*, i.e., B Th U 's available above $1,481^{\circ}\text{F}$ (805°C) per 1 lb of standard coal burnt (of 11 600 B Th U 's per lb) which occur in any given cement kiln with a given output

§ 8 Remarks on Table—Table I shows the gradual increase in the radiation losses as the kiln falls away from what is theoretically possible. The perfect kiln working with air preheated to $2,500^{\circ}\text{F}$ (the clinkering temperature) and producing 100 tons of clinker per 6 356 tons of standard coal burnt (or 11 73 lbs of clinker per 1 lb of coal burnt) shows no internal or external losses

As we go down the scale and come to the very imperfect kilns now used in actual practice, we see the internal radiation losses become perfectly enormous

E.g., a kiln producing 100 tons of clinker per 30 120 tons of coal has the enormous radiation loss of 4 922 B Th U 's per 1 lb of standard coal burnt when the air is not preheated at all, while if the air is preheated to 400°F , the loss is as great as 5 750 B Th U 's per 1 lb of standard coal burnt

§ 9 From this Table I we are now enabled to calculate the loss of clinker output due to internal radiation from any given rotary kiln provided that we know its output of clinker per 1 lb of standard coal (11 600 B Th U 's per lb) burnt and the external radiation losses from the clinkering and decarbonating zone

Suppose, for example, a given cement rotary kiln is yielding 100 tons of clinker per 30 120 tons of standard coal burnt. This is the same as 3 300 lbs of clinker per 1 lb of standard coal burnt

Let us suppose also that the entering air is preheated to 400°F which is the normal condition of the air entering most rotary kilns

$$\begin{aligned}\text{Then } Q - \text{B Th U 's required to heat 10 475 lbs of air from } 60^{\circ} \text{ to } 400^{\circ}\text{F} \\ &= 828 \text{ B Th U 's (Table I Chapter XII)} \\ 918 \text{ 6W heat absorbed in forming the clinker} \\ &= 918 \cdot 6 \times 3 \cdot 30 \\ &= 3 050 \text{ B Th U 's}\end{aligned}$$

Whence from equation (3)

$$\begin{aligned}R_i + R_e &= \text{total radiation losses} \\ 797 + 8 \cdot 8 &= 3050 \\ &= 5 750 \text{ B Th U 's per 1 lb of standard coal burnt}\end{aligned}$$

Also in Chapter XX it is shown that for an average modern rotary kiln the external radiation R_e from the decarbonating and clinkering zone is say, 469 B Th U 's

The internal radiation loss is

$$R_i = 5750 - 469 = 5281 \text{ B Th U 's per 1 lb of standard coal}$$

This represents a loss of high grade made *useless* for clinker production by being transformed into the low grade variety by radiation

The loss of clinker output thus occasioned is enormous

∴ Since every 918.6 B.Th.U.'s of high-grade heat loss represents the loss of 1 lb. of clinker, the loss due to 5,281 B.Th.U.'s is

$$\frac{5281}{918.6} = 5.75 \text{ lbs. per 1 lb. of standard coal burnt.}$$

In our present case 30.21 lbs. of coal produce 100 lbs. of clinker.

∴ On each 1 lb. of coal there is a loss in output of 5.75 lbs. of clinker.

∴ On 30.21 lbs. of coal there is a loss of $30.21 \times 5.75 = 174$ lbs. of clinker.

∴ If no internal radiation losses occurred, 30.21 lbs. of standard coal burnt would produce $174 + 100 = 274$ lbs. of clinker, or 100 lbs. of clinker would be produced by the expenditure of

$$\frac{30.21}{274} \times 100 = 11.0 \text{ lbs. of coal.}$$

In other words, the internal radiation prevailing inside a kiln normally producing 100 tons of clinker for 30.21 tons of standard coal of 12,600 B.Th.U.'s per 1 lb. is responsible for increasing its coal consumption from 11 to 30.21 tons of standard coal per 100 tons of clinker produced.

The internal radiation prevailing inside a cement rotary kiln is therefore the main factor in reducing the efficiency of the rotary kiln.

It is a far more effective factor than the external radiation, which, as we have seen in Chapter XX., is responsible for increasing its coal consumption from 26 to 30 tons of standard coal per 100 tons clinker produced.

The combined external and internal radiation amount to 5,750 B.Th.U.'s per 1 lb. of coal burnt in the case of the kiln previously considered (i.e., burning 30.120 lbs. of coal per 100 tons of clinker).

This total radiation means a loss of

$$\frac{5750}{918.6} = 6.26 \text{ lbs. of clinker per 1 lb. of coal.}$$

∴ The loss on 30.12 lbs. of coal is $30.12 \times 6.26 = 189$ lbs. of clinker.

Hence 30.12 tons of coal should be producing $100 + 189 = 289$ tons of clinker, or 100 tons of clinker would be produced by the expenditure of

$$\frac{30.12 \times 100}{289} = 10.4 \text{ tons of coal.}$$

So that the abolition of all radiation losses would reduce the fuel consumption of the kiln per 100 tons of clinker from 30.12 to only 10.4 tons of standard coal.

The following Table II. has been calculated in the above manner for different clinker outputs.

The table brings out the enormous losses which occur in the kiln due to radiation losses, and especially illustrates how much more serious are the internal radiation losses than the external.

Column (1) shows the fuel consumption of a rotary kiln expressed in tons of standard coal per 100 tons of clinker.

Column (2) shows what this fuel consumption would drop to if, by suitable insulation, all external radiation losses were completely stopped from the clinkering and decarbonating zone, but the internal radiation losses were left unchecked.

Column (3) shows what the fuel consumption would drop to if the internal radiation losses from the clinkering and decarbonating zone were completely stopped, but the external radiation losses were left unaltered.

Column (4) shows what the fuel consumption would become if both the internal and external radiation losses from the kiln were completely stopped.

TABLE II

Showing the Effect on Fuel Consumption of Stopping the Radiation Losses from Cement Kilns

| (1) Clinker Output from Kiln Tons Standard Coal Consumed per 100 Tons Clinker Made | (2) Effect of Stopping External Radiation from Clinkering and Decarbonating Zone but Keeping Internal Radiation Unaltered Tons Standard Coal Consumed per 100 Tons Clinker Made | (3) Internal Radiation from Clinkering and Decarbonating Zone Stopped but External Radiation Unaltered Tons Standard Coal Consumed per 100 Tons Clinker Made | (4) Both Internal and External Radiations Stopped in Clinkering and Decarbonating Zone Tons Standard Coal Consumed per 100 Tons Clinker Made |
|--|---|---|---|
| 20 855 | 18 86 | 11 02 | 10 44 |
| 25 044 | 22 22 | 11 02 | 10 44 |
| 26 021 | 22 97 | 11 02 | 10 44 |
| 27 078 | 23 79 | 11 02 | 10 44 |
| 28 209 | 24 73 | 11 02 | 10 44 |
| 29 137 | 25 36 | 11 02 | 10 44 |
| 30 170 | 26 11 | 11 02 | 10 44 |
| 34 305 | 29 19 | 11 02 | 10 44 |
| 39 246 | 32 70 | 11 02 | 10 44 |

Conclusions

(1) The radiation losses from the clinkering and decarbonating zone are extremely serious in increasing the coal consumption per 100 tons of clinker produced, the internal radiation losses being very much more deleterious than the external radiation losses in this respect

(2) The main cause of the thermal inefficiency of the rotary kiln is the enormous radiation losses

(3) In order to secure kiln economy the main line of progress for kiln designers lies in concentrating all efforts on eliminating as far as possible all internal and external radiation losses

Everything else exerts quite a subsidiary influence on kiln economy

CHAPTER XXII

EFFECT OF INTERNAL RADIATION AND CONVECTION ON SLURRY MOISTURE, SHOWING THAT WITHIN CERTAIN LIMITS THE PERCENTAGE OF MOISTURE IN A SLURRY MAY BE VARIED WITHOUT ALTERING THE FUEL CONSUMPTION PER 1 LB. OF CLINKER

§ 1. IN Chapter XXI. it was shown that in the ordinary cement rotary kiln as constructed to-day there occurs an enormous loss of high-grade heat in the upper part of the kiln, this high-grade heat being transferred by radiation or convection (by the hot gas) to the lower and colder regions of the kiln, and is there transformed into the equivalent amount of low-grade heat, without any useful chemical work being done in its passage from the high-grade to the low-grade state.

The nett result of this leakage of high-grade heat (*i.e.*, B.Th.U.'s available above $1,481^{\circ}$ F.) into the lower part of the kiln is that *more low-grade heat* (*i.e.*, B.Th.U.'s available below $1,481^{\circ}$ F.) *passes into the lower part of the kiln than is required for bringing the raw material to the correct temperature for entering the CO_2 expulsion zone and clinkering zone.*

§ 2. This excess of low-grade heat *cannot be remedied* except by improving the efficiency of the clinkering and decarbonating zones of the kiln, but of course it may be utilised in other ways—for example, in evaporating unnecessary water (*i.e.*, unnecessary for clinker formation) which has been placed in the slurry for convenience in manufacture.

§ 3. In a perfect kiln, of course, the slurry should be quite free from mechanically mixed water, so that all the heat available is directly used for producing clinker. In the imperfect practical kiln of to-day, where this heat is not so used, but leaks away into the lower region of the kiln, it will be possible to use this excess low-grade heat in boiling off excess water from the slurry *without increasing the fuel consumption per 1 ton of clinker formed.*

§ 4. If it was not so utilised, the wet slurry would find its way into the decarbonating zone and would absorb high-grade heat and thus seriously reduce the clinker output per 1 lb. of coal burnt, as explained in Chapter XV.

Hence *the advantage may be taken of the presence of excess of low-grade heat in the ordinary rotary kiln to add a certain amount of water to the slurry, which will be evaporated without affecting the clinker output per 1 lb. of coal* (which is decided by the quantity of high-grade heat available in the upper part of the kiln, and consequently by the design of the clinkering and decarbonating zones).

§ 5. If no excess water was present, the slurry being fed to the kiln in a dry state, all that would happen is that the *clinker output per 1 lb. of coal burnt*

would remain unchanged, but the temperature of the exit gases would increase. This is the reason why the *dry process* as practised in an ordinary cement rotary kiln does not yield the greater output to the extent expected of clinker per 1 lb of coal burnt than when the slurry is wet, provided that the slurry moisture does not exceed a certain percentage.

The chief advantage that the dry process would present would be that the exit gases from the dry process would issue at a higher temperature from the kiln than in the wet process, and therefore, if used in conjunction with a waste heat boiler, would generate more steam (but not much more clinker) per 1 lb of coal than is possible with the wet process*.

§ 6 Starting with a dry slurry, if we proceed to add successive quantities of water to the slurry, we will not alter much the yield of clinker per 1 lb of coal burnt until a certain critical percentage of water is reached (as shown at B in the diagram, Fig. 1), the exact value of B depending, in the case of each particular kiln, upon the efficiency in the design of the clinkering and decarbonating zones.

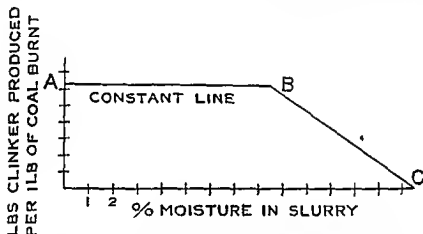


FIG. 1.—Diagram showing Effect of gradually increasing the Percentage of Moisture in a Slurry on the Number of Lbs. of Clinker produced per 1 Lb. of Coal burnt

When this percentage of water shown at B is reached then all the excess low grade heat (*i.e.* B Th U's available below 1481°F) in the lower part of the kiln is absorbed in evaporating the water and any excess of water beyond B will necessitate the utilisation of high grade heat in the upper part of the kiln for evaporating the water. This means that the high grade heat is withdrawn from the production of clinker and hence there will occur a *rapid decrease in the output of clinker per 1 lb of coal burnt* for each percentage of moisture beyond that indicated by B. The curve thus traverses the line BC. At C the production of clinker per 1 lb of coal burnt becomes *nil*, which means that the amount of water in the slurry is so great per 1 lb of coal burnt that the raw materials cannot be raised to the clinkering temperature.

In the ordinary rotary kiln the percentage of moisture used lies, as a rule, along *nc*, the point B being seldom reached.

In Chapters XIV and XV it was shown that to every 918.6 B Th U's of

* For simplicity we ignore the economising effect on the internal radiation losses due to the higher temperature of the colder end of the kiln. This was to have been treated in Part II of this Report but the premature dissolution of the British Portland Cement Research Association prevented the completion of this research.

high-grade heat (*i.e.*, B.Th.U.'s available above 1,481° F.) withdrawn in this way for evaporating water, there would occur a loss of 1 lb. of clinker formed.

§ 7. The following considerations may make the point clearer.

In Chapter IX. it was shown that in order to produce 1 lb. of clinker we must expend on the raw material 918.6 B.Th.U.'s above 1,481° F. (805° C.) and 652.5 B.Th.U.'s below 805° C. (1,481° F.).

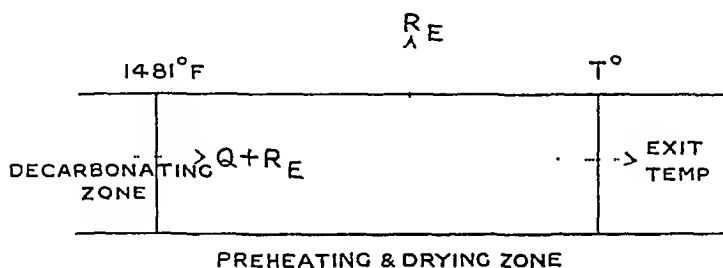


FIG. 2.

If the gases escape from the kiln at the exit temperature of T° F., and if Q be the total amount of heat which can be obtained from the gases corresponding to 1 lb. of clinker, escaping from the decarbonating zone and sinking to the exit temperature T° , and if R_i is the internal radiation (per 1 lb. of clinker produced) from the decarbonating zone into the dehydrating zone, and if R_E is the external radiation *from* the preheating and dehydrating zone, then obviously $Q + R_E - R_i - 652.5 =$ amount of heat available for heating X lbs. of water (per 1 lb. of clinker) from 60° to 212° F., and then superheating it to T° F.

So that

$$Q + R_E - R_i - 652.5 = X (212 - 60 + 970.7 + S[T - 212]),$$

$$\text{or} \quad X = \frac{Q + R_E - R_i - 652.5}{1122.7 + S(T - 212)} \quad (1)$$

where S is the mean specific heat of the steam between T° and 212° .

It is possible to calculate X knowing Q , R_E , and R_i . The calculations have been carried out in a number of cases, which will be given below.

Then, if the amount of water in the slurry is increased from 0 to X (per 1 lb. of clinker), the yield of clinker per 1 lb. of coal remains steady. When X is exceeded, the exit temperature begins to fall, until 212° F. is reached (below which it cannot go for practical purposes, as explained in a later chapter) and T becomes $= 212^\circ$ F.

So that (1) becomes

$$X = \frac{Q + R_E - R_i}{1122.7} \quad (2)$$

Provided Q , R_E , and R_i are not *altered*, any increase in X beyond this means that the yield of clinker per 1 lb. of coal begins to drop. So that X may vary from 0 to $\frac{(Q + R_E - R_i)}{1122.7}$ without affecting much the amount of clinker produced per 1 lb. of coal burnt.

Hence we can add to the slurry corresponding to 1 lb. of clinker a quantity of water which can range from 0 to X lbs. without affecting in the slightest the clinker output per 1 lb. of standard coal burnt.

Once, however, this limit of λ lbs is overstepped, it means that we have absorbed all the waste heat available, and consequently, in order to bring the raw material into a fit state (*i.e.*, heated to $1,481^{\circ}\text{F}$) for entering the decarbonating zone, we must use some of the high grade heat (*i.e.*, heat available above $1,481^{\circ}\text{F}$) for evaporating the water.

Each 918.6 B Th U's of high grade heat so absorbed means that 1 lb of clinker the less is produced in the upper zone. So that when the amount of moisture λ is exceeded, the clinker output begins to rapidly decrease, as above explained.

§ 8 It also follows from the preceding discussion that it is not possible to calculate the exact amount of water which can be present in a slurry without increasing the fuel consumption *unless the efficiency of the decarbonating and clinkering zone is previously determined*.

As we saw in § 7, the maximum value λ of the amount of water which can be added to that weight of slurry which will produce 1 lb of clinker is

$$\lambda = \frac{Q + R_2 - R_1}{1122.7}.$$

Q is at its minimum value when the gas escapes from the decarbonating zone at $1,481^{\circ}\text{F}$ (805°C), and which value can always be attained by lowering the speed with which the gas flows through the kiln.

R_1 (the internal radiation from the clinkering and decarbonating zone into the dehydrating zone) and R_2 (the external radiation from the dehydrating zone) are, however, fixed for a given type of kiln depending upon its shape, the thickness of the walls, etc. Hence in the main it is the factors R_2 and R_1 that decide the permissible value of λ .

Now R_1 is in the ordinary rotary kiln much greater than R_2 , so that R_1 is also the factor which in the main decides the efficiency of the decarbonating and clinkering zone. In other words the same factor R_1 which is predominant in deciding the efficiency of the clinkering and decarbonating zone is also predominant in deciding the amount of permissible water in the slurry. So that before λ can be calculated, R_1 must be known. The less is R_1 the more efficient is the decarbonising zone and the more we can reduce the percentage of water in the slurry and simultaneously reduce our fuel consumption per 1 lb of clinker formed.

§ 9 From the preceding discussion it will be seen that it is quite wrong to suppose that a reduction in the amount of water in the slurry will invariably and necessarily reduce the fuel consumption per 1 lb of clinker. One can vary the percentage of water in the fuel over quite a wide interval without affecting much the fuel consumption per 1 lb of clinker.

The extent of the possible variation depends upon the available excess of low grade heat (*i.e.*, heat available below $1,481^{\circ}\text{F}$) over 652.5 B Th U's per 1 lb of clinker, and this in its turn is decided by the efficiency of the clinkering and decarbonating zone in trapping and utilising the high grade heat (*i.e.*, B Th U's above $1,481^{\circ}\text{F}$) made available by the combustion of the coal.

It is, therefore, obvious that no possible alterations in the colder end of the kiln can effect a fuel economy in the upper end of the kiln (except as a secondary influence, which can be large, due to back radiation).

Once the high grade heat has escaped from the clinkering and decarbonating zone without doing any useful work in the hot part of the kiln, it is impossible to make it do this work afterwards, no matter what is done in the lower parts of the kiln.

The loss of heat must be stopped at its origin—that is, in the clinkering and decarbonating zone.

§ 10. Hence the introduction of lifter bars, etc., in the wet end of the kiln cannot improve directly the economy of the kiln *beyond those limits which are fixed by the efficiency of the clinkering and decarbonating zone.*

All that the lifter bars can do as regards fuel economy is to ensure that the low-grade heat is economically utilised in evaporating water in the slurry.

In cases where the introduction of lifter bars has been followed by a great improvement in the fuel economy, the effect is probably due to secondary causes—*e.g.*, by the checking of the too rapid flow of gas through the kiln, thus enabling the gas to give up its heat in the decarbonising zone. Also, by acting as a shield and causing an appreciable amount of back radiation, lifter bars, etc., may diminish the internal radiation from the decarbonating zone into the preheating zone.

To attempt to improve the economy of the kiln per 1 lb. of fuel by making the wet end more efficient without regarding the predetermined efficiency of the clinkering and decarbonating zone is just like shutting the stable door after the horse has escaped. It is vital to stop the high-grade heat escaping from the decarbonising zone into the lower part of the kiln. Once this has occurred, the process cannot be reversed, as follows from second law of thermodynamics. Once high-pressure water has lost its pressure in flowing from a high level to a low level, it has lost its power of doing work. So also, once high-temperature heat has lost its temperature in radiating from a high temperature to a low temperature, the heat has lost its power of making cement.

§ 11. We have just seen that it is not possible to calculate the exact amount of water which can be present in a slurry without increasing the fuel consumption unless the *efficiency of the clinkering and decarbonating zone is previously determined* (*i.e.*, unless the value of R_1 is known). What we can, however, do is this:—

If we assume that under the most favourable conditions the efficiency of the clinkering and decarbonating zone allowed a kiln to produce M lbs. of clinker per 1 lb. of standard coal burnt, then we can calculate quite easily how much water may be added to the slurry without the yield of the clinker per 1 lb. of coal burnt being in any way affected.

Also we can calculate what will be the exit temperature of the gases corresponding to any given quantity of moisture in the slurry. We will illustrate the process in the following chapters.

CHAPTER XXIII

CALCULATION OF THE AMOUNT OF WATER WHICH CAN BE RETAINED IN THE SLURRY SO AS TO ALLOW THE EXIT GASES TO ESCAPE FROM THE ROTARY KILN AT 212° F. (100° C.), ON THE SUPPOSITION THAT ALL EXTERNAL RADIATION FROM THE KILN SHELL IS STOPPED BY INSULATION

§ 1. THE reason for choosing a temperature of 212° F. for the exit gases is as follows:—

The cold slurry which enters a cement rotary kiln contains much mechanically mixed water, which must be heated by the outpouring hot combustion gases from the temperature at which it enters the kiln up to the boiling point (212° F. or 100° C.).

The water is then for the most part evaporated at 212° F., being turned into water vapour or steam. This steam is usually heated still higher than 212° F. by the hot gases coming from the interior of the kiln, and so in the ordinary kiln of to-day escapes in a superheated condition up the chimney. But now 212° F. (100° C.) is a very important theoretical temperature *inasmuch as it is the lowest temperature at which water can be expelled rapidly from the slurry*, and therefore 212° F. is the lowest practical temperature at which the exit gases can be *rapidly* removed from the kiln. We can imagine these exit gases being cooled and the water condensed afterwards, but *at the moment at which they escape from the kiln these gases must possess a temperature of at least 212° F.*, otherwise they could not rapidly carry away the moisture from the slurry. To expel the water *below 212° F. (100° C.)* from the slurry would either mean a long and slow drying process or else the use of a vacuum.

Hence we can assert *that 212° F. (100° C.) is the lowest possible practical temperature at which the exit gases can escape from a kiln, and once we have reduced the exit gases to this temperature, there lies little possibility of further substantial thermal economies being effected in the rotary kiln.*

§ 2. In Chapter XIX. a series of calculations were made as regards the temperatures at which the exit gases would emerge from a cement rotary kiln fulfilling certain conditions.

This exit temperature was shown in Table I., Chapter XIX., to remain steady at 212° F. when the yield of clinker ranged from 9.850 to 8.680 lbs. of clinker per 1 lb. of standard coal burnt, the temperature of the entering air needed for combustion of the coal being preheated from 500° to 60° F. by the outcoming clinker.

In these calculations the kiln was supposed to be correctly designed—that is all the heat was supposed to be trapped on the inside of the kiln and utilised in producing clinker instead of passing largely unutilised down the kiln and escaping up the chimney, as it does in the ordinary practical kiln of to day

§ 3 When the yield of clinker per 1 lb of coal burnt falls below theoretical figures given above and in Chapter XIV it means that a loss of high grade heat occurs in the upper part of the kiln this high grade heat being transferred by radiation and convection (by the hot gases) to the lower and colder regions of the kiln and is there transformed into the equivalent amount of low grade heat without doing any useful chemical work in its passage from the high grade to the low grade state

In Chapter XIV we dealt with the weight of clinker corresponding to a flame temperature of 3769°F In order to ensure continuity of treatment we will now take a series of values of clinker corresponding to a succession of lower flame temperatures say 3700°F In Chapter XIV it was shown that to every flame temperature there corresponded a definite weight of producible clinker and this is the order that will be followed here

The successive flame temperatures and the corresponding weight of clinker are set forth in Chapter XIV Table I and should be consulted by the reader

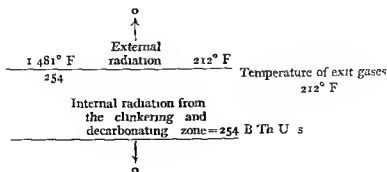
§ 4 Calculation of the maximum amount of water which can be retained in the slurry without increasing fuel consumption so as to allow the exit gases to escape at 212°F the air not being preheated

In Chapter XIV Table I it was shown that if 1 lb of standard coal united with 10.478 lbs of air to form 11.278 lbs of combustion gas at a uniform flame temperature of 3700°F then this 11.278 lbs of gas in falling from 3700°F to 1481°F can liberate an amount of heat which corresponds to the formation of 8.408 lbs of clinker per 1 lb of coal burnt or 11.893 tons of standard coal consumed per 100 tons of clinker produced

Corresponding to this weight of clinker are the following associated quantities (see Chapter XVII) —

| | | |
|----------------------------------|------------|--|
| CO_2 | 4.404 lbs | expelled at 1481°F (805°C) |
| Dry slurry | 13.119 lbs | |
| H_2O from kaolin | 0.2161 lb | expelled at 1472°F (800°C) |
| H_2O from silica | 0.0908 lb | expelled at 752°F (400°C) |

Also the number of B Th U s required to heat 13.119 lbs of slurry from 60° to 1481°F is weight of clinker $\times 652.5 = 5.486$ B Th U s (see Chapter IX)



By supposition the exit gas temperature is fixed at 212° F. The mean specific heat of the furnace gases between $1,481^{\circ}$ and 212° F. is 0.2610.

Also from Chapter XXI., Table I., we know that the sum of the internal and external radiation losses for the clinkering and decarbonating zone corresponding to data for this weight of clinker is 254 B.Th.U.'s when the air is not preheated.

Since the external radiation loss is taken as zero, the internal radiation loss is 254 B.Th.U.'s.

Hence: Heat lost by gas passing down preheating and decarbonating zone
 + Internal radiation into this zone from clinkering and decarbonating zone
 = Heat required to raise dry slurry to $1,481^{\circ}$ F.
 + Heat required to raise water contained in slurry from 60° to 212° F. and transform it into steam at 212° F.

Hence we get the following heat balance in the lower part of the kiln:—

(1) The B.Th.U.'s liberated by 11.278 lbs. of combustion gas sinking from $1,481^{\circ}$ to 212° F. are

$$11.278 \times 0.2610 \times (1481 - 212) = 3735.4 \text{ B.Th.U.'s.}$$

(2) Heat evolved from the 4.404 lbs. of CO_2 from the slurry sinking from $1,481^{\circ}$ F. (805° C.) to 212° F. (100° C.). From our tables this equals

$$4.404(360.911 - 37.483) = 1424.4 \text{ B.Th.U.'s.}$$

(3) Heat liberated when 0.2161 lb. of water vapour in the kaolin at $1,472^{\circ}$ F. (800° C.) sinks to 212° F. without condensing to liquid water. From our tables this will equal

$$0.2161(691.779 - 84.256) = 131.3 \text{ B.Th.U.'s.}$$

(4) The heat liberated when the 0.0908 lb. of water expelled from the hydrated silica at 752° F. (400° C.) sinks to 212° F. without condensing to water. This heat will be (from our tables)

$$0.0908(336.257 - 84.256) = 22.9 \text{ B.Th.U.'s.}$$

(5) Heat radiated from clinkering and decarbonating zone into the preheating and drying zone is 254 B.Th.U.'s. The quantities of heat in (1) and (4) must equal:—

(a) The amount of heat required to raise 13.119 lbs. of dry slurry from 60° to $1,481^{\circ}$ F. preparatory to its entering the decarbonating zone. This will absorb 5,486 B.Th.U.'s (see Chapter IX.), being the weight of clinker $\times 652.5$.

(b) The amount of heat required to raise x lbs. of water mechanically mixed with the 13.119 lbs. of slurry from 60° to 212° F., and turn it into vapour at 212° F. This amounts to

$$x \times (212 - 60 + 970.7) = x \times 1122.7 \text{ B.Th.U.'s.}$$

(c) The amount of heat lost in external radiation in the present case is assumed to be zero.

Hence we obtain the following equation:—

$$3735.4 + 1424.4 + 131.3 + 22.9 + 254 = 5486 + 1122.7x + 0 \quad (1)$$

or

$$x = 0.073 \text{ lb. of water.}$$

This means that under the circumstances specified, the 13 119 lbs of dry slurry may have admixed with it 0.073 lb of water, and still the exit gases will escape at 212° F

But 13 119 lbs of dry slurry + 0.073 moisture = 13 192 wet slurry, or 13 192 lbs wet slurry may contain 0.073 lb water, or 100 000 lbs wet slurry may contain 0.554 lb of water, which can be raised to the boil from 60° F and expelled as steam at 212° F

§ 5 Modification of the Preceding Calculation when the Entering Air is Preheated to Various Amounts—In the preceding calculation it has been assumed that the air has not been preheated by the issuing clinker. Under these circumstances, with the given clinker output, the given radiation loss would be 254 B Th U's, as given in Table I, Chapter XXI

If the air is preheated to different degrees and yet the clinker output per 1 lb of standard coal burnt remains unaffected, the radiation losses must be altered. Thus the following shows the radiation losses (taken from Table I, Chapter XXI) —

| | | | | |
|---|-------|--------|---------|---------|
| <i>Output of clinker per 1 lb of standard coal burnt, 8 408 lbs</i> | | | | |
| <i>Temperature of preheated air</i> | 60° F | 400° F | 2039° F | 2500° F |
| <i>Corresponding internal radiation losses in B Th U's</i> | 254 | 1082 | 5414.4 | 6733 |

If the preceding calculation be repeated and, instead of putting the internal radiation losses as = 254 B Th U's, the other figures from the table of Chapter XXI are substituted, we get the possible allowance of water for the cases when the air is preheated to various other temperatures

E.g., if the entering air is preheated to 400° F, the internal radiation loss is 1,082 B Th U's, and we get instead of equation (1)

$$3735.4 + 1424.4 + 131.3 + 22.9 + 1082 = 5486 + 1122.7x$$

or

$$x = 0.8106 \text{ lb of water}$$

So that when the air is preheated to 400° F 13 119 lbs of dry slurry may have admixed with it 0.811 lb of water or 13 119 + 0.811 = 13.93 lbs of wet slurry may contain 0.811 lb of water

Whence 100 lbs may contain $\frac{0.811}{13.93} \times 100 = 5.83$ lbs of water, which could be

heated from 60° F to the boiling point and evaporated at 212° F

Proceeding in this way for different yields of clinker per 1 lb of coal and for various degrees of preheating of the entering air, the table on pp 236 and 237 has been calculated

§ 6 Remarks on the Table—This table is very instructive as regards showing the increase of fuel consumption with an increasing percentage of moisture in the slurry, and especially as regards showing the importance of preheating the entering air as highly as possible if it is desired to reduce the fuel consumption to the lowest possible limit

It shows, for example, what will happen when all external radiation from a kiln is stopped

(a) Consider, for example, the case when the air entering the kiln is not preheated at all (i.e., enters at 60° F), the results are set forth in column IV of Table I

If the kiln output is to be maintained at 100 tons of clinker per 11.52 tons of standard coal, the exit gases being kept at 212° F. (100° C.), then no water at all can be mixed with the slurry.

On adding water to the dry slurry so as to make the percentage of moisture in the wet slurry 0.554 per cent., we find that this will increase the fuel consumption to 11.893 tons of standard coal per 100 tons of clinker (still keeping the exit gases at 212° F.). An increase in the slurry moisture to 4.522, 24.87, 36.96, and 55.07 per cent. will cause the corresponding fuel consumption to increase to 12.535, 16.770, 20.855, and 30.12 tons of standard coal per 100 tons of clinker, respectively.

(b) Now consider the case when the entering air is preheated to 400° F., the exit temperature of the gases being kept steady at 212° F. (100° C.). This case is shown in column V. of the table. Here no water at all can be present in the slurry if the fuel consumption is to be maintained at 10.74 tons of standard coal per 100 tons of clinker.

If water be added to the slurry to bring the percentage up to 1.43 per cent., then the fuel consumption will increase to 11.0558 tons of standard coal per 100 tons of clinker. Further increase in succession of the slurry moisture to 9.62, 40.63, and 57.76 per cent. will cause the fuel consumption to increase respectively to 12.525, 20.855, and 30.120 tons of standard coal per 100 tons of clinker.

(c) Columns VI. and VII. show the effect of arranging the kiln so that *all* the heat in the *outcoming clinker* is utilised in preheating the incoming air (the latter being kept at 10.478 lbs. per 1 lb. of standard coal). Then, if we maintained the exit gases at 212° F., we see that 20.52 per cent. of moisture in the slurry would cause a fuel consumption of 10.4384 tons of standard coal per 100 tons of clinker. 30.58 per cent. of moisture would increase the fuel consumption to 13.210 tons of coal per 100 tons of clinker, where 40.50 per cent. of moisture would cause the fuel consumption to increase to 16.770 lbs., while 61.18 per cent. of slurry moisture would increase the fuel consumption to 30.120 tons of fuel per 100 tons of clinker.

(d) Finally, column VIII. shows the effect of heating all the incoming air to the clinkering temperature of $2,500^{\circ}$ F. before letting it enter the combustion zone.

Under these circumstances a fuel consumption of 10.1526 tons of standard coal per 100 tons of clinker is possible with a slurry moisture content of 20.90, while a fuel consumption of 30.120 tons of standard coal per 100 tons of clinker would allow of a slurry moisture of 70.06 per cent.

§ 7. The most important point brought out by Table I., however, is the fact that it is *not the moisture in the slurry that is instrumental in causing the poor clinker output per lb. of coal burnt in the modern rotary kiln.*

Thus a glance at the table will show that with the entering air preheated to 400° F. (which is what may be taken as normal for a modern kiln), the slurry moisture could be 40.63 per cent. and the fuel consumption would only be 20.855 lbs. of standard coal per 100 tons of clinker.

Seeing that the average modern kiln working with a slurry of this composition consumes, say, 33 tons of standard coal per 100 tons of clinker, the enormous difference between what the practical kiln consumes and what it should consume is certainly not due to the moisture content of the slurry, but must be due to other causes, such as the ineffective and bad design of the clinkering and decarbonating zones.

The efforts made in recent years to improve the fuel economy of rotary kilns by concentrating on reducing the water content of the slurry are largely due to a faulty grasp of the theoretical thermodynamical principles which govern the kiln.

The same efforts applied to improve the efficiency of the clinkering and decarbonating zones, especially as regards utilising the heat of the outcoming

TABLE

*Showing the Amount of Water which can be Retained in Slurry so as to
All External Radiation from*

| Air Supply 10 478 Lbs. per 1 Lb Standard Coal | | | Maximum Percentage Slurry Moisture which Enables an Exit Temperature of 212° F to be Maintained with Air Entering the Kiln Preheated to the Following Temperatures | | | | |
|--|--|---|---|--|--|-------------------------|--|
| Clinker Production | | Tempera- ture of Exit Gases | Air at 60° F Corre- sponding Slurry Moisture | Air at 400° F Corre- sponding Slurry Moisture | Air at Maximum Temperature that it can be Heated to by Outgoing Clinker * | | Air at 2,300° F Corre- sponding Slurry Moisture |
| Tons of Standard Coal per 100 Tons Clinker | Lbs. of Clinker Produced per 1 Lb Standard Coal | | | | Corre- sponding Slurry Moisture | Tempera- ture of Air | |
| I | II | III | IV | V | VI | VII | VIII |
| 6 3565 | 15 732 | ° F 60 (cannot be kept at 212° F) | Per Cent | Per Cent | Per Cent | ° F (2500) | Per Cent |
| 10 1526 | 9 850 | 212 | | | 19 32 | (2352) | 20 90 |
| 10 4384 | 9 580 | 212 | | | 20 52 | (2293) | 21 72 |
| 10 7400 | 9 311 | 212 | | | 21 78 | (2236) | 24 54 |
| 11 0558 | 9 045 | 212 | | 1 43 | 23 05 | (2178) | 26 37 |
| 11 3830 | 8 785 | 212 | | 2 43 | 24 32 | (2121) | 27 34 |
| 11 5207 | 8 680 | 212 | | 3 36 | 24 88 | (2102) | 29 90 |
| 11 893 | 8 408 | 212 | 0 56 | 5 83 | 26 24 | (2039) | 31 63 |
| 12 525 | 7 984 | 212 | 4 52 | 9 62 | 28 43 | (1944) | 33 8 |
| 13 210 | 7 570 | 212 | 3 52 | 13 45 | 30 58 | (1853) | 36 8 |
| 13 957 | 7 165 | 212 | | | 33 03 | (1764) | |
| 14 791 | 6 761 | 212 | | | 35 44 | (1674) | |
| 15 726 | 6 359 | 212 | | | 37 95 | (1584) | |
| 16 770 | 5 963 | 212 | 24 87 | 29 09 | 40 50 | (1494) | 46 75 |
| 17 960 | 5 568 | 212 | | | | | |
| 19 301 | 5 181 | 212 | | | | | |
| 20 855 | 4 795 | 212 | 36 96 | 40 63 | 48 38 | (1226) | 57 58 |
| 30 120 | 3 320 | 212 | 55 07 | 57 76 | 61 18 | (881) | 70 06 |

* There are 10 478 lbs. of air at 60° F entering the kiln per 1 lb. of standard coal burnt, then be capable, under the most favourable conditions of heating the air to the temperatures see Chapter XVI

I

Allow the Exit Gases to Escape from the Rotary Kiln at 212° F. (100° C.), the Kiln Shell being Stopped

Radiation Losses per 1 Lb. Standard Coal Burnt (12,600 B.Th.U.'s per Lb.).

| Internal Radiation from Clinkering and Decarbonating Zone into Preheating and Dehydrating Zone. | | | | | External Radiation from Kiln Shell. | |
|---|---|--|-----------------------|---|---|--|
| Air at 60° F. Corresponding Internal Radiation. B.Th.U.'s per 1 Lb. Standard Coal. | Air at 400° F. Corresponding Internal Radiation. B.Th.U.'s per 1 Lb. Standard Coal. | Air at Maximum Temperature that it can be Heated to by Out-coming Clinker (see Column VII.). Corresponding Internal Radiation. | | Air at 2,500° F. Corresponding Internal Radiation. B.Th.U.'s per 1 Lb. Standard Coal. | From Pre-heating and Dehydrating Zone to External Air. B.Th.U.'s per 1 Lb. Standard Coal. | From Clinkering and Decarbonating Zone and End of Kiln. B.Th.U.'s per 1 Lb. Standard Coal. |
| | | B.Th.U.'s per 1 Lb. Standard Coal. XI. | Temperature Air. XII. | | | |
| IX. | X. | | | XIII. | XIV. | XV. |
| ... | ... | ... | ° F. (2500) | ... | ... | ... |
| ... | ... | 4972.0 | (2352) | 5,403 | ... | ... |
| ... | ... | 5052.0 | (2293) | 5,651 | ... | ... |
| ... | 247 | 5137.0 | (2236) | 5,898 | ... | ... |
| ... | 491 | 5216.0 | (2178) | 6,142 | ... | ... |
| ... | 730 | 5294.0 | (2121) | 6,381 | ... | ... |
| ... | 828 | 5338.3 | (2102) | 6,479 | ... | ... |
| 254 | 1082 | 5414.4 | (2039) | 6,733 | ... | ... |
| 638 | 1466 | 5532.0 | (1944) | 7,117 | ... | ... |
| 1018 | 1846 | ... | ... | ... | ... | ... |
| 1390 | 2218 | ... | ... | ... | ... | ... |
| 1792 | 2590 | ... | ... | ... | ... | ... |
| 2131 | 2959 | ... | ... | ... | ... | ... |
| 2494 | 3322 | 6151.0 | (1494) | 8,973 | ... | ... |
| 2857 | ... | ... | ... | ... | ... | ... |
| 3213 | ... | ... | ... | ... | ... | ... |
| 3567 | 4395 | 6509.0 | (1226) | 10,046 | ... | ... |
| 4922 | 5750 | 6963.0 | (881) | 11,401 | ... | ... |

while there are X lbs. of clinker (shown in column II.) issuing at 2,500° F. This clinker will shown in column VI. For list of temperatures corresponding to different weights of clinker,

clinker to preheat the entering air, could not fail to have produced much greater economies

As shown above, the real cause of the inefficiency is the loss of high grade heat in the upper part of the kiln (due to radiation both internal and external, and to convection by the hot gases), which escapes unabsorbed down into the lower part of the kiln and is wasted in evaporating water or in raising the temperature of the exit gases

CHAPTER XXIV

CALCULATION OF THE AMOUNT OF WATER WHICH CAN BE RETAINED IN THE SLURRY SO AS TO ALLOW THE EXIT GASES TO ESCAPE AT 212° F., ALLOWING FOR THE ORDINARY EXTERNAL RADIATION LOSSES WHICH OCCUR IN A MODERN ROTARY KILN

§ 1. In Chapter XXIII. we calculated the amount of water which could be retained in the slurry to give a given clinker output per lb. of coal burnt, the exit gases being kept at the lowest possible temperature capable for getting rid of the steam, viz., 212° F.

In these calculations it was assumed that the kilns were so carefully insulated that all external radiation from the kiln shell was stopped, so that all that we had to contend with was the internal radiation.

In the following chapter similar calculations are carried out, but in them the kiln is supposed to be losing heat by external radiation to the same amount as occurs in actual practice, as shown by kiln measurements carried out by the British Portland Cement Research Association during recent years.

§ 2. It is now necessary to decide on the allowance to be made for radiation. In Chapter XX. the radiation losses from the different zones of a modern rotary kiln were analysed and were classified as follows:—

| | | B.Th.U.'s per 1 Lb. Standard Coal of 12,600 B.Th.U.'s Burnt. |
|--|-----------|---|
| (1) From end sections of kiln | | 84 |
| (2) In the clinkering zone | | 210 |
| (3) In the CO_2 expulsion zone | | 259 |
| (4) In the dehydrating and preheating zone | | 147 |
| | | 553 |

We may take it, then, that from the hot upper portion of the kiln—above the preheating and dehydrating zone—we lose 553 B.Th.U.'s, below 147 B.Th.U.'s, for every 12,600 B.Th.U.'s liberated in the kiln when 1 lb. of standard coal is burnt.

These are the figures which will be assumed in the following calculations

§ 3. We must now explain the method of performing the calculations. For this purpose we will take a typical case as an illustration. In Chapter XIV. it was shown that corresponding to a flame temperature of $3,500^{\circ}$ F. one could produce 7.570 lbs. of clinker per 1 lb. of standard coal burnt. In Chapter XVIII. it was shown that the following quantities were associated with this weight of clinker:—

Weight of dry slurry required to produce 7.570 lbs. of clinker is 11.811 lbs.

Weight of CO_2 expelled from the slurry at $1,481^{\circ}$ F. (805° C.) is 3.965 lbs.

Weight of water expelled from the kaolin at $1,472^{\circ}$ F. (800° C.) is 0.1945 lb.

Weight of water expelled from the silica at 752° F. (400° C.) is 0.0818 lb.

The number of B Th U 's required to raise the slurry required to make 7 570 lbs of clinker from 60° to $1,481^{\circ}$ F—the temperature of the decarbonating zone—is $7\,570 \times 652.5 = 4939.0$ B Th U 's (see Chapter IX)

The exit gas temperature is assumed to be 212° F, and the problem now before us is to calculate the maximum amount of water which may be added to the slurry to maintain the exit temperature at this figure. This we will now proceed to do

- (1) The amount of heat liberated by 11 278 lbs of combustion gas sinking from $1,481^{\circ}$ to 212° F without any steam condensing is 3735.4 B Th U 's (see Chapter XIX, § 4 (a))
- (2) The amount of heat liberated by 3 965 lbs of CO_2 from the slurry sinking from $1,481^{\circ}$ to 212° F is (from our heat tables) $3\,965 \times 323.428 = 1282.4$ B Th U 's
- (3) The amount of heat liberated when 0 1945 lb of water vapour expelled from the kaolin at $1,472^{\circ}$ F sinks to 212° F without condensing to liquid water is $0.1945 \times 607.523 = 118.2$ B Th U 's
- (4) The heat liberated when the 0 0818 lb of water expelled from the hydrated silica at 752° F (400° C) sinks to 212° F without condensing to water will be $0.0818 \times 252.001 = 20.6$ B Th U 's
- (5) Heat brought into clinkering zone by preheated air per 1 lb of coal burnt is

| Temperature Air $^{\circ}$ F | B Th U 's Carried in |
|---------------------------------|-------------------------|
| 60 | 0 0 |
| 400 | 828 0 |
| 1853 | 4642 0 |
| 2500 | 6479 0 |

- (6) From the table of internal radiation, given in Chapter XXI, we get external + internal radiation from clinkering and decarbonating zone into preheating and drying zone per 1 lb of coal burnt —

| Temperature Air $^{\circ}$ F | Total Radiation |
|---------------------------------|-------------------------------|
| 60 | $1018 + 0 = 1018.0$ B Th U 's |
| 400 | $1018 + 828 = 1846.0$ |
| 1853 | $1018 + 4642 = 5660.0$ |
| 2500 | $1018 + 6479 = 7497.0$ |

- (7) The external radiation from clinkering and decarbonating zone and end of kiln per 1 lb of coal burnt is 553.0 B Th U 's (assumed as normal for a modern kiln)
- (8) The external radiation from preheating and drying zone per 1 lb of coal burnt is 147.0 B Th U 's (assumed as normal for a modern kiln)
- (9) The internal radiation from the clinkering and decarbonating zone into the preheating and drying zone per 1 lb of coal burnt is

| Entering Air $^{\circ}$ F | Internal and External Radiation | External Radiation | Internal Radiation |
|------------------------------|------------------------------------|-----------------------|-----------------------|
| 60 | 1018 | 553 = | 465.0 B Th U 's |
| 400 | 1846 | 553 = | 1293.0 „ |
| 1853 | 5660 | 553 = | 5107.0 „ |
| 2500 | 7497 | 553 = | 6944.0 „ |

We then have the following equation connecting these quantities (calculated on per 1 lb of clinker) —

Quantity of heat lost by gases in sinking from $1,481^{\circ}$ to 212° F + quantity of heat radiated (or convected) into the preheating and dehydrating zone

=quantity of heat required to raise the dry slurry from 60° to $1,491^{\circ}$ F. + quantity of heat required to raise X lbs. of water mixed with the dry slurry from 66° to 212° F. and then turn it into steam at 212° F.

Equating these quantities, we obtain the following series of equations:—

(a) *Entering Air at 60° F.*—

$$3735.4 + 1282.4 + 118.2 + 20.6 + 465 = 4939.0 + 1122.7X + 147.0,$$

whence

$$X = 0.4771 \text{ lb. water.}$$

Hence 11.811 lbs. of dry slurry may contain 0.4771 lb. H_2O ; or 11.811 + 0.4771 = 12.288 lbs. wet slurry may contain 0.4771 lb. H_2O ; or 100 lbs. wet slurry may contain 3.883 lbs. H_2O .

(b) *Entering Air at 400° F.*—Here the internal radiation is 1,293 B.Th.U.'s instead of 465. Performing the same calculation as before, we get $X = 1.215$ lbs. of water, or 100 lbs. of wet slurry may contain 9.326 lbs. H_2O .

(c) *Entering Air at $1,853^{\circ}$ F.*—Here internal radiation is 5,107 B.Th.U.'s, and we get $X = 4.612$ lbs. of water, or 100 lbs. of wet slurry may contain 28.08 lbs. of water.

(d) *Air at $2,500^{\circ}$ F.*—Here internal radiation is 6,944 B.Th.U.'s, and we get $X = 6.248$ lbs. of water, or 100 lbs. of wet slurry may contain 34.60 lbs. of water.

§ 4. By repeating this calculation for all the weights of clinker given in Table I., column (3), Chapter XIV., and the associated weights of dry slurry, carbon dioxide, etc., given in Chapter XVIII., Table I., we get a series of values which are set forth in Table I.

§ 5. **Comments on the Table.**—A study of this table shows that in general the remarks made on the similar table in the preceding chapter can be applied to this table as well, the external radiation being not sufficient in amount to effect the character of the data. In an average rotary kiln the entering air may be taken as being preheated to 400° F. By looking at column V. of the table it will be seen that if a rotary kiln be fed with slurry containing 38.24 per cent. of moisture and the exit gases escaped at 212° F., the fuel consumption of a properly designed kiln should be only 20.855 tons of standard coal per 100 tons clinker, against 30 to 33 tons in the ordinary kiln.

A kiln consuming 30.120 tons of standard coal could bear a slurry moisture as high as 52.40 per cent.

It is obvious, therefore, that it is not the amount of moisture in the slurry which is the effective agent in increasing the fuel consumption of a kiln to the neighbourhood of, say, 30 to 33 tons of standard coal per 100 tons of clinker, but some other cause is at work. This, in the preceding chapters, has been traced to the leakage of high-grade heat (*i.e.*, B.Th.U.'s available above $1,481^{\circ}$ F.) from the hot part of the kiln to the colder parts without performing useful work in its passage.

Hence improvements in the efficiency of the rotary kiln must be looked for *not so much in reducing the moisture in the slurry as in increasing the efficiency of the kiln in the hotter regions.*

§ 6. **Limit of Waste-heat Boiler.**—Another point which is brought out by the table is the limits of the possible use of the waste-heat boiler, the kiln being supposed to lose heat by external radiation at the ordinary rate prevalent in a rotary kiln at the present time. Unless the exit gases can escape at a higher temperature than 212° F., they cannot generate steam at a technically useful pressure when passed through a boiler.

Since the table gives the maximum amount of water in the slurry which just allows the gases to escape at 212° F., we can see at once what limits are set to

Showing the Amount of Water which can be Retained in Slurry so as (100° C), Allowing the Same External Radiation from

| Air Supply 10 478 Lbs per 1 Lb Standard Coal of 12,600 B Th U's | | | Maximum Percentage Slurry Moisture which Enables an Exit Temperature of 212° F to be Maintained with Air Entering the Kiln Preheated to the Following Temperatures | | | | |
|---|--|--------------------------------------|---|--|---|----------------------------|--------------------|
| Clinker Production | | Tempera- ture of Exit Gases | Air at 60° F Corre- sponding Slurry Moisture | Air at 400° F Corre- sponding Slurry Moisture | Air at Maximum Temperature that it can be Heated to by Outcoming Clinker * | | Air at 2,500° F |
| Tons of Standard Coal per 100 Tons Clinker | Lbs of Clinker per 1 Lb Standard Coal Burnt | | | | Corre- sponding Slurry Moisture | Tempera- ture of Air | |
| | | | | | | | |
| | | ° F | Per Cent | Per Cent | Per Cent | ° F | Per Cent |
| 11 893 | 8 408 | 212 | 0 000 | 1 40 | 23 57 | (2039) | 28 46 |
| 12 525 | 7 984 | 212 | 0 000 | 5 35 | 25 77 | (1944) | 31 53 |
| 13 210 | 7 570 | 212 | 3 883 | 9 326 | 28 08 | (1853) | 34 60 |
| 13 957 | 7 165 | 212 | 8 03 | 13 28 | 30 43 | (1764) | 37 64 |
| 14 791 | 6 761 | 212 | 12 27 | 17 33 | 32 89 | (1674) | 40 72 |
| 15 726 | 6 359 | 212 | 16 57 | 21 43 | 35 43 | (1584) | 43 83 |
| 16 770 | 5 963 | 212 | 20 89 | 25 50 | 38 05 | (1494) | 46 92 |
| 17 960 | 5 563 | 212 | 25 31 | 29 76 | 40 78 | (1403) | 50 08 |
| 19 301 | 5 181 | 212 | 29 73 | 33 97 | 43 61 | (1315) | 53 21 |
| 20 855 | 4 795 | 212 | 34 24 | 38 24 | 46 55 | (1226) | 56 37 |
| 22 660 | 4 413 | 212 | 38 81 | 42 57 | 49 60 | (1137) | 59 55 |
| 24 801 | 4 032 | 212 | 43 46 | 46 97 | 52 80 | (1047) | 62 77 |
| 25 044 | 3 993 | 212 | 43 95 | 47 44 | 53 16 | (1041) | 63 10 |
| 25 278 | 3 956 | 212 | 44 40 | 47 83 | 53 49 | (1033) | 63 41 |
| 25 517 | 3 919 | 212 | 44 87 | 48 30 | 53 81 | (1024) | 63 74 |
| 25 760 | 3 882 | 212 | 45 30 | 48 70 | 54 10 | (1015) | 64 00 |
| 26 021 | 3 843 | 212 | 45 80 | 49 20 | 54 50 | (1006) | 64 20 |
| 26 274 | 3 806 | 212 | 46 30 | 49 60 | 54 80 | (997) | 64 70 |
| 26 532 | 3 769 | 212 | 46 90 | 50 10 | 55 00 | (988) | 65 00 |
| 26 802 | 3 731 | 212 | 47 20 | 50 20 | 55 20 | (979) | 65 20 |
| 27 078 | 3 693 | 212 | 47 70 | 51 00 | 55 80 | (970) | 65 70 |
| 27 352 | 3 656 | 212 | 48 20 | 51 40 | 56 15 | (962) | 66 00 |
| 27 632 | 3 619 | 212 | 48 60 | 51 80 | 56 50 | (953) | 66 30 |
| 27 917 | 3 582 | 212 | 49 00 | 52 30 | 56 80 | (944) | 66 60 |
| 28 209 | 3 545 | 212 | 49 60 | 52 70 | 57 20 | (935) | 67 00 |
| 28 514 | 3 507 | 212 | 50 00 | 53 20 | 57 50 | (926) | 67 30 |
| 28 827 | 3 469 | 212 | 50 50 | 53 70 | 57 90 | (917) | 67 60 |
| 29 137 | 3 432 | 212 | 51 00 | 54 10 | 58 20 | (910) | 67 90 |
| 29 453 | 3 395 | 212 | 51 50 | 54 60 | 58 50 | (899) | 68 30 |
| 29 771 | 3 359 | 212 | 51 90 | 55 00 | 58 90 | (891) | 68 60 |
| 30 120 | 3 320 | 212 | 52 40 | 55 50 | 59 20 | (881) | 68 90 |
| 30 460 | 3 283 | 212 | 52 90 | 55 90 | 59 60 | (872) | 69 20 |
| 34 305 | 2 915 | 212 | 57 70 | 60 40 | 63 20 | (781) | 72 50 |
| 39 246 | 2 549 | 212 | 62 60 | 65 00 | 66 90 | (694) | 75 80 |
| 45 766 | 2 185 | 212 | 67 16 | 69 70 | 70 90 | (605) | 79 10 |
| 54 824 | 1 824 | 212 | 72 60 | 74 40 | 75 00 | (516) | 82 40 |
| 70 126 | 1 466 | 212 | 77 80 | 79 30 | 79 40 | (428) | 85 80 |
| 90 000 | 1 111 | 212 | 82 90 | 84 10 | 81 90 | (340) | 89 10 |
| 131 579 | 0 760 | 212 | 88 20 | 89 00 | 88 70 | (252) | 92 50 |
| 243 309 | 0 413 | 212 | 93 60 | 94 00 | 93 70 | (164) | 95 90 |
| 1538 459 | 0 065 | 212 | 99 00 | 99 00 | 99 00 | (76) | 99 40 |
| Infinite | 0 000 | 212 | 100 00 | 100 00 | 100 00 | (60) | 100 00 |

* There are 10 478 lbs of air at 60° F entering the kiln per 1 lb of standard coal burnt, then be capable, under the most favourable conditions of heating the air to the temperatures Chapter XVI

I

to Allow the Exit Gases to Escape from the Rotary Kiln at 212° F.
the Kiln Shell as Occurs in a Modern Rotary Kiln

| Radiation Losses per 1 Lb. Standard Coal Burnt (12,600 B.Th.U.'s per Lb.). | | | | | | |
|---|--|--|-----------------------|---|---|--|
| Internal Radiation from Clinkering and Decarbonating Zone into Preheating and Dehydrating Zone. | | | | External Radiation from Kiln Shell. | | |
| Air at 60° F. Corresponding Internal Radiation. B.Th.U.'s per 1 Lb. Standard Coal. IX. | Air at 400° F. Corresponding Internal Radiation. B.Th.U.'s per 1 Lb. Standard Coal. X. | Air at Maximum Temperature that it can be Heated by Outcoming Clinker (see Col. VII.). Corresponding Internal Radiation. | | Air at 2,500° F. Corresponding Internal Radiation. B.Th.U.'s per 1 Lb. Standard Coal. XIII. | From Preheating and Dehydrating Zone and End of Kiln to External Air. B.Th.U.'s per 1 Lb. Standard Coal. XIV. | From Clinkering and Decarbonating Zone and End of Kiln. B.Th.U.'s per 1 Lb. Standard Coal. XV. |
| | | B.Th.U.'s per 1 Lb. Standard Coal. XI. | Temperature Air. XII. | | | |
| | | | ° F. | | | |
| ... | 529 | 4861 | (2039) | 6,180 | 553 | 147 |
| 85 | 913 | 4979 | (1944) | 6,564 | 553 | 147 |
| 465 | 1293 | 5107 | (1853) | 6,944 | 553 | 147 |
| 837 | 1665 | 5234 | (1764) | 7,316 | 553 | 147 |
| 1209 | 2037 | 5356 | (1674) | 7,688 | 553 | 147 |
| 1578 | 2406 | 5479 | (1584) | 8,057 | 553 | 147 |
| 1941 | 2769 | 5598 | (1494) | 8,420 | 553 | 147 |
| 2304 | 3132 | 5716 | (1403) | 8,783 | 553 | 147 |
| 2660 | 3488 | 5837 | (1315) | 9,139 | 553 | 147 |
| 3014 | 3842 | 5956 | (1226) | 9,493 | 553 | 147 |
| 3366 | 4194 | 6073 | (1137) | 9,845 | 553 | 147 |
| 3715 | 4543 | 6186 | (1047) | 10,194 | 553 | 147 |
| 3751 | 4579 | 6207 | (1041) | 10,230 | 553 | 147 |
| 3785 | 4613 | 6220 | (1033) | 10,264 | 553 | 147 |
| 3819 | 4647 | 6230 | (1024) | 10,298 | 553 | 147 |
| 3853 | 4681 | 6241 | (1015) | 10,333 | 553 | 147 |
| 3889 | 4717 | 6254 | (1006) | 10,368 | 553 | 147 |
| 3923 | 4751 | 6264 | (997) | 10,402 | 553 | 147 |
| 3957 | 4785 | 6275 | (988) | 10,436 | 553 | 147 |
| 3992 | 4820 | 6287 | (979) | 10,471 | 553 | 147 |
| 4027 | 4855 | 6298 | (970) | 10,506 | 553 | 147 |
| 4061 | 4889 | 6312 | (962) | 10,540 | 553 | 147 |
| 4095 | 4923 | 6322 | (953) | 10,574 | 553 | 147 |
| 4129 | 4957 | 6333 | (944) | 10,608 | 553 | 147 |
| 4163 | 4991 | 6344 | (935) | 10,642 | 553 | 147 |
| 4198 | 5022 | 6355 | (926) | 10,677 | 553 | 147 |
| 4232 | 5060 | 6366 | (917) | 10,711 | 553 | 147 |
| 4266 | 5094 | 6382 | (910) | 10,745 | 553 | 147 |
| 4300 | 5128 | 6387 | (899) | 10,779 | 553 | 147 |
| 4334 | 5162 | 6401 | (891) | 10,813 | 553 | 147 |
| 4369 | 5197 | 6410 | (881) | 10,848 | 553 | 147 |
| 4403 | 5231 | 6421 | (872) | 10,882 | 553 | 147 |
| 4742 | 5570 | 6528 | (781) | 11,221 | 553 | 147 |
| 5078 | 5906 | 6642 | (694) | 11,557 | 553 | 147 |
| 5412 | 6240 | 6752 | (605) | 11,891 | 553 | 147 |
| 5743 | 6571 | 6859 | (516) | 12,222 | 553 | 147 |
| 6072 | 6900 | 6969 | (428) | 12,551 | 553 | 147 |
| 6398 | 7226 | 7078 | (340) | 12,877 | 553 | 147 |
| 6721 | 7549 | 7186 | (252) | 13,200 | 553 | 147 |
| 7041 | 7869 | 7292 | (164) | 13,520 | 553 | 147 |
| 7359 | 8187 | 7446 | (76) | 13,838 | 553 | 147 |
| 7972 | 8800 | 7972 | (60) | 14,451 | 553 | 147 |

while there are X lbs. of clinker (shown in column II.) issuing at 2,500° F. This clinker will be shown in column VI. For list of temperatures corresponding to different weights of clinker, see

the use of a waste heat boiler when the entering air is preheated to different amounts

For example, a kiln with a fuel consumption of 11.893 tons of standard coal per 100 tons of clinker could not have any water in the slurry if the air was preheated to only 60° F. If the air was preheated to 400° F, however, a limit to the moisture in the slurry would be 1.40 per cent. If the slurry contained *less* moisture than this, a little steam could be generated.

If, however, the entering air be preheated to the maximum possible amount by the outgoing clinker, the slurry could contain 23.57 per cent of moisture, and yet the gases could escape at 212° F, so that this slurry moisture sets a limit to the employment of the waste heat boiler here. The slurry moisture must be *below* 23.57 per cent before the waste heat boiler could be used.

In the ordinary rotary kiln as it exists at the present time, we may consider that the slurry moisture is about 40 per cent and the entering air is preheated to 400° F. Looking at column V of the table, we see that when the slurry moisture is 38.24 per cent and the air enters at 400° F, the fuel consumption would be 20.855 tons of standard coal if the exit gases escaped at 212° F. So that under practical conditions a kiln working with a fuel consumption of 21 tons of coal per 100 tons of clinker and a slurry moisture of 40 per cent could not under any circumstances permit of the application of a waste heat boiler as the temperature of the exit gases would be too low.

If the amount of water in the slurry is reduced below 40 per cent (other things being held equal), the exit temperature could increase above 212° F, and under such conditions an extension of the range over which the waste heat boiler can be applied is possible. Even here, however, the range of possible extension is not great, as will be seen later.

CHAPTER XXV

CALCULATION OF THE EXIT TEMPERATURES OF THE GASES FROM ROTARY KILNS WHEN THE SLURRY MOISTURE VARIES BETWEEN 0 AND 40 PER CENT., BUT THE ENTERING AIR IS PREHEATED TO VARIOUS DEGREES, ALLOWING THE SAME EXTERNAL RADIATION FROM KILN SHELL AS OCCURS IN AN ORDINARY ROTARY KILN

§ 1. It is important to calculate the exit temperatures of the gases issuing from cement rotary kilns when they use a slurry containing different percentages of moisture.

Such a series of calculations have been carried out below for slurries containing 40, 30, 20, 10, and 0 per cent. of moisture, the entering air being supposed to be preheated to different amounts, as set forth in Tables III., IV., V., VI., and VII. below.

§ 2. In order to explain the method of calculation, a typical case will be worked through.

The figures we will work through will range from 4.413 to 1.466 lbs. of clinker per 1 lb. of coal, with a slurry containing 40 per cent. of moisture.

We will work out in full a typical calculation for the case of 4.413 lbs. of clinker, and give in the form of a table the data for the other quantities.

§ 3. In the following typical case the amount of clinker produced per 1 lb. of standard coal burnt is taken as 4.413 lbs., corresponding to a flame temperature of 2,700° F. (see Chapter XIV., Table I.).

Then we have the following associated quantities (see Chapter XVII., Table I.) per 1 lb. of standard coal burnt in the kiln :—

Clinker, 4.413 lbs.

Slurry, 6.886 lbs.

CO_2 expelled from slurry is 2.312 lbs.

H_2O expelled from kaolin at 1,472° F. (800° C.) is 0.1134 lb.

H_2O expelled from silica at 752° F. (400° C.) is 0.0477 lb.

B.Th.U.'s required to heat 6.886 lbs. of slurry from 60° to 1,481° F. are weight of clinker $\times 652.5 = 2,880$ B.Th.U.'s. The wet *slurry* contains 40 per cent. of moisture.

It is required to calculate the exit temperatures of the issuing gases when the 10.478 lbs. of entering air are preheated to various temperatures by the issuing gases.

§ 4 We have now the following relationship —

- (1) 60 lbs of dry slurry are mixed with 40 lbs water in 100 lbs wet slurry

Hence 6886 lbs of dry slurry are mixed with 4591 lbs of water

- (2) The exit gas temperature $T^{\circ}\text{F}$ is known by a preliminary calculation to lie between 300° and 212°F

- (3) The mean specific heat of the gases of combustion between $1,481^{\circ}$ and 300°F is 0.2624. The mean specific heat of the combustion gas between 212° and 300°F is, from our tables (see Chapter VI)

$$\frac{0.2421 + 0.2450}{2} = 0.24355$$

- (4) The B Th U's liberated by 11278 lbs of combustion gas sinking (a) from $1,481^{\circ}$ to 300°F , (b) from 300° to $T^{\circ}\text{F}$ are

$$11278 \times 0.2624 \times 1181 + 11278 \times 0.24355(300 - T) \\ = 34945 + 27468(300 - T)$$

- (5) From our tables the instantaneous specific heat of CO_2 at 300°F is 0.2219 and the instantaneous specific heat of CO_2 at 200°F is 0.2144

Hence the mean specific heat of CO_2 between 300° and 200°F is $\frac{1}{2}(0.2219 + 0.2144) = 0.21815$

- (6) The B Th U's liberated by 2312 lbs of CO_2 sinking from $1,481^{\circ}$ to 300°F , and then from 300° to $T^{\circ}\text{F}$ are, from our tables

$$2312(360.911 - 56.912) + 2312 \times 0.21815(300 - T) \\ = 7028 + 0.5044(300 - T)$$

- (7) The instantaneous specific heat of water vapour at 300°F is 0.4665 and at 200°F is 0.4670

Hence the mean specific heat between 200° and 300°F is 0.46675

- (8) The B Th U's liberated when 0.1134 lb of water expelled from the kaolin at 1472°F sinks first to 300°F and then from 300° to $T^{\circ}\text{F}$ are

$$0.1134(691.679 - 175.323) + 0.1134 \times 0.46675(300 - T) \\ = 64.2 + 0.0529(300 - T)$$

- (9) The B Th U's liberated when the 0.0477 lb of water expelled from the hydrated silica at 75°F (a) sinks to 300°F , (d) and then from 300° to $T^{\circ}\text{F}$ are

$$0.0477(336.257 - 125.323) + 0.0477 \times 0.46675(300 - T) \\ = 10.1 + 0.0223(300 - T)$$

- (10) The B Th U's required to raise 4591 lbs of water in slurry from 60° to 212°F , gasify it at 212°F , and heat the steam from 212° to $T^{\circ}\text{F}$ are

$$4591 \times 1122.7 + 4591 \times 0.46675(T - 212) \\ = 51543 + 21428(T - 212)$$

- (11) The internal radiation from the clinkering and decarbonating zone into the preheating and drying zone per 1 lb of coal burnt is given by the following table —

TABLE I

| Temperature Entering Air. ° F | Radiation in B.Th.U.'s. |
|-------------------------------------|----------------------------|
| 60 | 3366 |
| 100 | 3366 + 96.5 |
| 200 | 3366 + 338.0 |
| 300 | 3366 + 582.0 |
| 400 | 3366 + 828.0 |
| 500 | 3366 + 1076.0 |
| 600 | 3366 + 1327.0 |
| 700 | 3366 + 1579.0 |
| 800 | 3366 + 1834.0 |
| 900 | 3366 + 2090.0 |
| 1000 | 3366 + 2349.0 |
| 1137 | 3366 + 2707.0 |
| 2500 | 3366 + 6479.0 |

(12) The external radiation from the preheating and drying zone per 1 lb. of coal burnt is taken as 147 B.Th.U.'s.

(13) Hence we have the following series of equations for determining the exit temperatures T :—

(a) Case where air enters at 60° F.

$$3494.5 + 2.7468(300 - T) + 702.8 + 0.5044(300 - T) + 64.2 + 0.0529(300 - T) + 10.1 + 0.0223(300 - T) + 3366.0 = 2880 + 147 + 5154.3 + 2.1428(T - 212),$$

whence

$$T = \frac{908.4}{5.4692} = 166.2^\circ \text{ F.}$$

(b) Case where air enters at 100° F.

Every quantity will be the same in the above equation, except that the internal radiation will have increased from 3366 to 3366 + 96.5 (see (11) above). This will increase T to

$$T = \frac{908.4 + 96.5}{5.4692} = 183.7^\circ \text{ F.}$$

(c) Case where air enters at 200° F.

Here everything will be the same, except that the internal radiation will have increased from 3366 to 3366 + 338, and hence

$$T = \frac{908.4 + 338}{5.4692} = 227.9^\circ \text{ F.}$$

(d) Air entering at 300° F.

$$T = \frac{908.4 + 582}{5.4692} = 272.5^\circ \text{ F.}$$

(e) Air entering at 400° F.

$$T = \frac{908.4 + 828}{5.4692} = 317.4^\circ \text{ F.}$$

(f) Air entering at 500° F.

$$T = \frac{908.4 + 1076}{5.4692} = 362.8^\circ \text{ F.}$$

(g) Air entering at 600° F.

$$T = \frac{908.4 + 1327}{5.4692} = 408.8^\circ \text{ F.}$$

(h) Air entering at 700° F.

$$T = \frac{908.4 + 1579}{54692} = 454.8^\circ \text{ F.}$$

(i) Air entering at 800° F.

$$T = \frac{908.4 + 1834.0}{54692} = 501.5^\circ \text{ F}$$

(j) Air entering at 900° F.

$$T = \frac{908.4 + 2090}{54692} = 548.2^\circ \text{ F.}$$

(k) Air entering at 1,000° F.

$$T = \frac{908.4 + 2349}{54692} = 595.5^\circ \text{ F.}$$

(l) Air entering at 1,137° F.

$$T = \frac{908.4 + 2707}{54692} = 661^\circ \text{ F.}$$

(m) Air entering at 2,500° F.

$$T = \frac{908.4 + 6479}{54692} = 1,351^\circ \text{ F}$$

§ 5 By repeating this calculation for each of the various amounts of clinker contained in Chapter XIV, Table I, column (3), and repeated in column (2) of the table on p 255, the following Table III results for a slurry containing 40 per cent of moisture

Tables IV, V, VI, and VII for slurries containing successively 30, 20, 10, and 0 per cent of moisture are calculated in precisely the same manner from the same data, allowing for the diminished amounts of moisture in the slurry

§ 6 By interpolation the exit temperatures appertaining to any other degree of moisture can be easily calculated from the results given in Tables III, IV, V, VI, and VII. For example, a slurry is used containing 45 per cent of moisture. What will be the exit temperature when the entering air is preheated to 400° F if the kiln is consuming 28.209 tons of standard coal per 100 tons of clinker produced?

Referring to Tables III, IV, V, VI, and VII below, and looking up the data appertaining to a coal consumption of 28.209 tons per 100 tons clinker with the entering air preheated to 400° F (which is about what is normal in an ordinary kiln), we get the table on p 255

We can now proceed in several different ways in order to obtain the required result

For example, by plotting the slurry moisture against the exit temperatures we get a continuous curve from which may be read off the exit temperature corresponding to any given slurry moisture

For the particular case of 45 per cent of moisture, the result, as read off from the curve, is 560° F.

It is more accurate to calculate by interpolation. For example, referring to Table II, an increase in the slurry moisture by 10 per cent (viz, from 30 to 40 per cent) lowers the exit temperature by 405.4° F (viz, from 1169.0° to 763.6° F.), hence an increase in slurry moisture by 5 per cent would lower the exit temperature by $\frac{5}{10} \times 405.4 = 202.7^\circ \text{ F}$. So that the exit temperature corresponding to 45 per cent moisture would be $763.6^\circ - 202.7^\circ \text{ F} = 560.9^\circ \text{ F}$.

TABLE II

| (1) Slurry Moisture. | (2) Exit Temperature. | (3) 1st Difference. Δ_1 | (4) 2nd Difference. Δ_2 | (5) 3rd Difference. Δ_3 | (6) 4th Difference. |
|----------------------------|-----------------------------|--------------------------------------|--------------------------------------|--------------------------------------|------------------------|
| | ° F. | ° F. | ° F. | ° F. | ° F. |
| 0 | 2293.8 | | | | |
| 10 | 1937.8 | -356.3 | | | |
| 20 | 1561.0 | -376.5 | -20.0 | +4.7 | |
| 30 | 1169.0 | -392.0 | -15.5 | +2.1 | -2.6 |
| 40 | 763.6 | -405.4 | -13.4 | | |

This method is accurate enough for all practical purposes. Should, however, even greater accuracy be desired, we can use Newton's interpolation formula, and so obtain any degree of approximation desired.

The general formula is *

$$f(a+xw) = f(a) + x \cdot \Delta_1 + \frac{x \cdot (x-1)}{1 \cdot 2} \Delta_2 + \frac{x \cdot (x-1) \cdot (x-2)}{1 \cdot 2 \cdot 3} \Delta_3 + \frac{x \cdot (x-1)(x-2)(x-3)}{1 \cdot 2 \cdot 3 \cdot 4} \Delta_4 + \dots$$

Hence, referring to our Table II.

$$a = 40 \text{ per cent.}, w = 10, a+xw = 45; \therefore xw = 5,$$

or $x = 0.5$. $\Delta_1, \Delta_2, \Delta_3$, etc., are the 1st, 2nd, and 3rd differences.

$$f(a) = 763.6^\circ \text{ F.}$$

Substituting in our formula

$$\begin{aligned} f(45^\circ) &= 763.6^\circ + 0.5 \times (-405.4) + \frac{(0.5)(0.5-1)}{1 \cdot 2} (-13.4) \\ &\quad + \frac{(0.5)(0.5-1)(0.5-2)}{1 \cdot 2 \cdot 3} (+2.1) \\ &\quad + \frac{(0.5)(0.5-1)(0.5-2)(0.5-3)}{1 \cdot 2 \cdot 3 \cdot 4} (-2.6) \\ &= 763.60 - 202.70 + 1.67 + 0.13 + 0.10 \\ &= 562.8^\circ \text{ F.} \end{aligned}$$

Now, take a slightly more difficult case. Suppose that we wish to calculate the exit temperature corresponding to 43.7 per cent. moisture, all other conditions being the same as in the previous example. Then

$$a = 40, w = 10, f(a+x \cdot w) = f(43.7), \text{ and since } a+xw = 43.7, \therefore xw = 3.7,$$

or $x = 0.377$.

* See Whittaker, "A Short Course in Interpolation," 1923, pp. 5, 11.

Hence, substituting in the formula

$$\begin{aligned}
 f(43.7) &= 763.6 + 0.377(-405.4) + \frac{(0.377)(0.377-1)}{1^2} (-13.4) \\
 &\quad + \frac{(0.377)(0.377-1)(0.377-2)}{1^2 \cdot 2} (+2.1) \\
 &\quad + \frac{(0.377)(0.377-1)(0.377-2)(0.377-3)}{1^2 \cdot 2 \cdot 3 \cdot 4} (-2.6) \\
 &= 763.60 - 152.83 + 1.57 + 0.13 + 0.23 \\
 &= 612.7^\circ \text{F}
 \end{aligned}$$

If only the 1st differences were used the figure would become $763.60 - 152.83 = 610.77^\circ \text{F}$, which is close enough for practical purposes.

It will therefore be seen that the given tables afford enough data to make any necessary calculation regarding the exit temperature appertaining to any required slurry moisture with a kiln possessing any given coal consumption per 100 tons clinker.

§ 7 Comments on Table III for 40 per Cent Slurry Moisture—It will be noticed that the exit temperature obtained corresponding to a kiln burning 30.12 tons of standard coal per 100 tons of clinker produced is 891.7°F if the air is preheated to 400°F , and 792.7°F if the air is only preheated to 200°F . These correspond closely to the exit temperatures observed in practice.

The fact that the exit temperatures so calculated do in fact correspond to those obtained in actual practice shows that the theory underlying the calculations is correct, and that therefore any deductions made from that theory should have due weight attached to them.

Now, very important deductions as regards losses of heat by internal and external radiation are based upon the theory.

So that the results obtained in this chapter furnish an additional proof of the correctness of the causes deduced above for the inefficiency of the modern rotary kiln.

These causes were, it will be remembered, the excessive internal radiation from the hotter part of the kiln to the colder parts, the moisture in the slurry playing only a subordinate part in the loss of economy.

Hence the moral pointed out by the table is that the correct way to abolish inefficiency of the kiln is to concentrate on making the hot end of the kiln most efficient and not to worry unduly about the slurry moisture.

§ 8 Comments on Tables III, IV, V, VI, and VII—It will be noticed that the main effect of reducing the slurry moisture, keeping other factors the same, is to increase the exit temperatures of the issuing gases.

Thus a kiln consuming 78.206 tons of standard coal per 100 tons clinker, with entering air preheated to 400°F , will give an exit temperature of 764°F with a 40 per cent slurry moisture, but if the moisture is cut down to 30 per cent, the exit temperature will rise to $1,169^\circ \text{F}$, and if the slurry is fed in dry, the exit temperature will rise to $2,294^\circ \text{F}$, as set forth in the series of tables.

§ 9 Hence the main effect of reducing the moisture in the slurry is to increase the amount of steam producible by the exit gases when the latter are passed through a waste heat boiler.

The amount of clinker produced per 1 lb. of coal burnt depends so largely upon the efficiency of the hot end of the kiln that in general a reduction in the amount of moisture in the slurry will not produce the expected increase in the amount of clinker.

The amount of clinker produced per 1 lb. of coal burnt depends entirely (as we have seen) upon the number of *B.Th.U.*'s available in the hot gases above $1,481^\circ \text{F}$, and not at all upon the amount available below, so that the main factor governing the clinker output is the efficiency of the hot end of the kiln.

§ 10. The tables also indicate the utmost limit of application of a waste-heat boiler to a kiln fed with slurries containing different percentages of moisture.

It is obvious, for example, that before a waste-heat boiler can be utilised the exit gases must escape at a higher temperature than 212° F. (100° C.)—the boiling-point of water.

Now, by looking at the successive tables, we can at once see the limits of application of the boiler. Thus, referring to Table III. for 40 per cent. slurry moisture, we see that if we could so improve the efficiency of the hot end of the kiln that it consumes 20.885 lbs. of standard coal per 100 tons clinker, the entering air being preheated to 400° F., then the exit temperature is only 142.3° F.—not high enough to raise any steam.

So that it will be impossible to fit usefully a waste-heat boiler to a kiln fed with 40 per cent. slurry when its coal consumption drops to about 21 tons of standard coal per 100 tons of clinker. Here the entering air is preheated to 400° F. If the clinker gave up all its heat to the entering air (see column XIV., Table III.), which enters preheated to nearly $1,500^{\circ}$ F. (see column XV., Table III.), so that the kiln was working in an ideal manner, the limit would be shifted to between 16.770 and 17.960 tons of standard coal per 100 tons clinker, for the exit temperature would, in the first case, be 126.7° F., and in the second case 246.8° F.—too low for exit heat boilers.

The case is somewhat more favourable when the slurry moisture is reduced.

Thus, assuming that the entering air is preheated to 400° F., we see by Table IV. that when the slurry moisture is 30 per cent. the exit temperature is at 217.7° F. when the fuel consumption is 17.960 tons of standard coal per 100 tons clinker; with 20 per cent. slurry moisture the corresponding figures are 265.6° F. for a fuel consumption of 15.726 tons of standard coal.

For 10 per cent. moisture, 336.1° F. for 13.957 tons of standard coal, and for a perfectly dry kiln (0 per cent. moisture) we get an exit temperature of 344.7° F. for a fuel consumption as low as 11.893 tons of standard coal per 100 tons of clinker.

Hence there is a considerable prospect of extending the use of the waste-heat boiler by reducing the slurry moisture. The limits of the use of the waste-heat boiler are marked on the tables.

§ 11. **Upper and Lower Higher Limit of Exit Temperatures.**—It will be noticed that the calculations of exit temperatures are not given below that appertaining to a kiln consuming 70.126 tons of standard coal per 100 tons of clinker produced (*i.e.*, 1.466 lbs. of clinker per 1 lb. of standard coal burnt in kiln).

The reason is that these temperatures are impossible to realise in practice and for the following reason.

In order that clinker should be produced at all, a clinkering temperature of at least $2,000^{\circ}$ F. is necessary, and even then formation of clinker at these temperatures will require a very long time.

By turning to Chapter XIV. we see that a low efficiency such as would yield clinker at approximately the above rates will correspond to a flame temperature in the clinkering zone lower than that at which clinker can be produced. So that when the inefficiency of the kiln falls below a certain value, no clinker at all is produced. It is certain that a kiln burning 70 tons of coal per ton of clinker would probably only turn out underburnt clinker, and therefore any deductions regarding the exit temperatures of such a kiln are imaginary.

In other words, just as there exists a definite upper limit to kiln efficiency (6.36 tons of standard coal per 100 tons clinker), so also there exists a definite lower limit to kiln efficiency, the last being reached when the inefficiency of the hot zone is so poor that the clinkering temperature is not attained.

Note.—These tables are calculated from interpolated values of specific heats, and therefore can only be regarded as sufficiently accurate for works engineers.

TABLE

Showing the Exit Temperatures of the Gases Issuing from a Rotary Kila when
 is Preheated to Various Degrees
 Air Supply—10 478 lbs. per 1 lb of standard coal of 12,600 B Th U's

| Clinker Production | | Entering Air at 60° F Corre- sponding Exit Temp | Entering Air at 100° F Correspond- ing Exit Temp | Entering Air at 200° F Correspond- ing Exit Temp | Entering Air at 300° F Correspond- ing Exit Temp | Entering Air at 400° F Correspond- ing Exit Temp | Entering Air at 500° F Correspond- ing Exit Temp |
|--|---|---|---|---|---|---|---|
| Tons of Standard Coal per 100 Tons of Clinker | Lbs of Clinker per 1 Lb of Standard Coal | | | | | | |
| I | II | III | IV | V | VI | VII | VIII |
| | | ° F | ° F | ° F | ° F | ° F | ° F |
| 11 893 | 8 408 | | | | | | |
| 12 525 | 7 984 | | | | | | |
| 13 210 | 7 570 | | | | | | |
| 13 957 | 7 165 | | | | | | |
| 14 791 | 6 761 | | | | | | |
| 15 726 | 6 359 | | | | | | |
| 16 770 | 5 963 | | | | | | |
| 17 960 | 5 568 | | | | | | |
| 19 301 | 5 181 | | | | | | |
| 20 855 | 4 795 | | | 56 4 | 99 0 | 142 3 | 185 8 |
| 22 660 | 4 413 | 166 2 | 183 7 | 327 9 | 272 5 | 317 4 | 362 8 |
| 24 801 | 4 032 | 349 5 | 367 8 | 413 7 | 461 1 | 506 9 | 554 1 |
| 25 044 | 3 993 | 368 6 | 387 0 | 432 8 | 479 2 | 526 0 | 573 1 |
| 25 278 | 3 956 | 387 0 | 405 5 | 451 6 | 498 3 | 545 3 | 592 7 |
| 25 517 | 3 919 | 405 5 | 424 0 | 470 3 | 517 1 | 564 2 | 612 4 |
| 25 760 | 3 882 | 425 4 | 444 0 | 490 5 | 537 5 | 584 8 | 632 5 |
| 26 021 | 3 843 | 445 3 | 464 0 | 510 7 | 557 9 | 605 5 | 654 4 |
| 26 274 | 3 806 | 464 7 | 483 5 | 530 4 | 577 8 | 625 5 | 673 8 |
| 26 532 | 3 769 | 482 9 | 501 7 | 548 5 | 595 9 | 643 7 | 691 8 |
| 26 802 | 3 731 | 500 8 | 519 6 | 566 4 | 613 7 | 661 4 | 709 6 |
| 27 078 | 3 693 | 522 5 | 541 3 | 588 2 | 635 8 | 683 8 | 732 1 |
| 27 352 | 3 656 | 539 9 | 558 6 | 605 6 | 653 2 | 700 8 | 749 3 |
| 27 632 | 3 619 | 559 7 | 578 6 | 625 9 | 673 6 | 721 6 | 770 2 |
| 27 917 | 3 582 | 580 0 | 598 9 | 646 3 | 694 2 | 742 6 | 791 2 |
| 28 209 | 3 545 | 600 2 | 619 2 | 666 9 | 715 0 | 763 6 | 812 5 |
| 28 514 | 3 507 | 621 4 | 640 5 | 688 4 | 736 7 | 785 6 | 834 8 |
| 28 827 | 3 460 | 642 4 | 661 6 | 709 7 | 758 3 | 807 4 | 856 7 |
| 29 137 | 3 432 | 663 4 | 682 8 | 731 1 | 780 0 | 829 2 | 879 0 |
| 29 455 | 3 395 | 681 5 | 700 8 | 749 1 | 797 9 | 847 0 | 896 6 |
| 29 771 | 3 359 | 702 3 | 721 7 | 770 2 | 819 2 | 868 6 | 918 5 |
| 30 120 | 3 320 | 724 7 | 744 2 | 792 9 | 842 2 | 891 7 | 941 9 |
| 30 460 | 3 283 | 746 4 | 766 0 | 815 0 | 864 4 | 914 3 | 964 5 |
| 34 305 | 2 915 | 889 8 | 910 0 | 960 0 | 1010 0 | 1061 0 | 1112 0 |
| 39 246 | 2 549 | 1218 0 | 1240 0 | 1293 0 | 1347 0 | 1402 0 | 1457 0 |
| 45 766 | 2 185 | 1483 5 | 1506 0 | 1562 0 | 1619 0 | 1676 3 | 1734 0 |
| 54 824 | 1 824 | 1740 0 | 1763 0 | 1821 0 | 1880 0 | 1939 0 | 1999 0 |
| 70 126 | 1 466 | 2063 0 | 2088 0 | 2149 0 | 2211 0 | 2273 0 | 2337 0 |

Limits of Waste Heat Boiler

III

the Slurry Moisture is Retained Constant at 40 per Cent., but the Entering Air by the Outgoing Clinker

per lb. entering at 60° F. Clinkering temperature assumed to be 2,500° F.

| Entering Air at 600° F. Corresponding Exit Temp. | Entering Air at 700° F. Corresponding Exit Temp. | Entering Air at 800° F. Corresponding Exit Temp. | Entering Air at 900° F. Corresponding Exit Temp. | Entering Air at 1,000° F. Corresponding Exit Temp. | Entering Air at Maximum Temp. to which it can be Preheated by Outcoming Clinker. | | Entering Air at 2,500° F. Corresponding Exit Temp. |
|--|--|--|--|--|--|-----------------------|--|
| | | | | | Corresponding Exit Temp. of Gas. | Maximum Temp. of Air. | |
| IX. | X. | XI. | XII. | XIII. | XIV. | XV. | XVI. |
| ° F. | ° F. | ° F. | ° F. | ° F. | ° F. | ° F. | ° F. |
| ... | ... | ... | ... | ... | ... | ... | ... |
| ... | ... | ... | ... | ... | ... | ... | ... |
| ... | ... | ... | ... | ... | ... | ... | ... |
| ... | ... | ... | ... | ... | ... | ... | 101.2 |
| ... | ... | ... | ... | ... | ... | ... | 246.6 |
| ... | ... | ... | ... | ... | ... | 1584 | 401.5 |
| ... | ... | ... | ... | ... | 126.7 | 1494 | 566.0 |
| ... | ... | ... | 32.9 | 74.8 | 246.8 | 1403 | 743.0 |
| 63.5 | 105.9 | 148.9 | 191.9 | 235.5 | 374.8 | 1315 | 930.5 |
| 229.8 | 274.0 | 318.7 | 363.5 | 408.9 | 524.6 | 1226 | 1133.0 |
| 408.8 | 454.8 | 501.5 | 548.2 | 595.4 | 661.0 | 1137 | 1351.0 |
| 601.7 | 649.7 | 698.1 | 746.0 | 796.0 | 819.1 | 1047 | 1581.0 |
| 620.7 | 668.6 | 717.1 | 765.7 | 814.9 | 835.3 | 1041 | 1599.0 |
| 640.7 | 688.8 | 737.6 | 786.5 | 836.6 | 852.4 | 1031 | 1625.0 |
| 660.5 | 708.8 | 757.7 | 806.8 | 856.4 | 867.6 | 1024 | 1647.2 |
| 680.8 | 729.3 | 778.4 | 827.7 | 877.6 | 885.2 | 1015 | 1673.0 |
| 701.9 | 750.7 | 800.0 | 849.5 | 899.6 | 902.8 | 1006 | 1698.0 |
| 722.6 | 771.5 | 821.1 | 870.8 | 921.1 | 919.5 | 997 | 1723.0 |
| 740.5 | 789.4 | 838.9 | 888.6 | 938.9 | 932.9 | 988 | 1741.0 |
| 758.3 | 807.2 | 856.6 | 906.3 | 956.5 | 946.0 | 979 | 1757.6 |
| 781.0 | 830.1 | 879.8 | 929.7 | 980.1 | 965.0 | 970 | 1784.9 |
| 798.1 | 847.2 | 896.8 | 946.7 | 997.1 | 977.9 | 962 | 1801.0 |
| 819.3 | 868.6 | 918.5 | 968.5 | 1019.2 | 1005.1 | 953 | 1827.0 |
| 840.5 | 890.0 | 940.1 | 990.4 | 1041.2 | 1012.9 | 944 | 1852.7 |
| 862.1 | 911.8 | 962.1 | 1012.6 | 1063.7 | 1031.0 | 935 | 1879.0 |
| 884.5 | 934.5 | 985.0 | 1036.0 | 1087.0 | 1049.1 | 926 | 1906.0 |
| 906.8 | 957.0 | 1008.0 | 1059.0 | 1110.0 | 1068.0 | 917 | 1933.0 |
| 929.1 | 979.6 | 1030.0 | 1082.0 | 1134.0 | 1086.0 | 910 | 1960.2 |
| 946.9 | 997.2 | 1048.0 | 1099.0 | 1151.0 | 1099.0 | 899 | 1977.0 |
| 968.9 | 1020.0 | 1071.0 | 1122.0 | 1174.0 | 1117.4 | 891 | 2003.5 |
| 992.5 | 1043.0 | 1095.0 | 1147.0 | 1199.0 | 1136.5 | 881 | 2032.0 |
| 1015.0 | 1066.0 | 1118.0 | 1170.0 | 1223.0 | 1156.0 | 872 | 2060.0 |
| 1164.0 | 1217.0 | 1270.0 | 1323.0 | 1376.0 | 1260.0 | 781 | 2232.0 |
| 1513.0 | 1568.0 | 1625.0 | 1682.0 | 1739.0 | 1565.2 | 694 | 2654.6 |
| 1792.0 | 1851.0 | 1910.0 | 1970.0 | 2030.0 | 1795.4 | 605 | 2990.8 |
| 2060.0 | 2120.0 | 2182.0 | 2243.0 | 2304.0 | 2008.0 | 516 | 3299.0 |
| 2400.0 | 2464.0 | 2524.0 | 2594.0 | 2660.0 | 2291.0 | 428 | 3709.0 |

TABLE

*Showing the Exit Temperatures of the Gases Issuing from a Rotary Kilm when
is Preheated to Various Degrees*

Air Supply — 10 478 lbs per 1 lb of standard coal of 12,600 B Th U s

| Clinker Production | | Entering Air at 60° F Corre- sponding Exit Temp | Entering Air at 100° F Correspond- ing Exit Temp | Entering Air at 200° F Correspond- ing Exit Temp | Entering Air at 300° F Correspond- ing Exit Temp | Entering Air at 400° F Correspond- ing Exit Temp | Entering Air at 500° F Correspond- ing Exit Temp |
|--|---|---|---|---|---|---|---|
| Tons of Standard Coal per 100 Tons of Clinker | Lbs of Clinker per 1 Lb of Standard Coal | III | IV | V | VI | VII | VIII |
| I | II | ° F | ° F | ° F | ° F | ° F | ° F |
| 11 893 | 8 408 | | | | | | |
| 12 525 | 7 984 | | | | | | |
| 13 210 | 7 570 | | | | | | |
| 13 957 | 7 165 | | | | | | |
| 14 791 | 6 761 | | | | | | |
| 15 726 | 6 359 | Limits of Waste Heat Boiler | | | | | |
| 16 770 | 5 963 | | | | | 46 9 (imposs) | 92 9 |
| 17 960 | 5 568 | 44 3 (imposs) | 62 8 | 109 1 | 155 9 | 217 7 | 250 6 |
| 19 301 | 5 181 | 202 5 | 221 6 | 269 5 | 317 9 | 366 6 | 415 8 |
| 20 855 | 4 795 | 374 2 | 396 6 | 443 6 | 493 8 | 544 4 | 595 3 |
| 22 660 | 4 413 | 550 1 | 570 6 | 622 1 | 673 8 | 726 2 | 709 4 |
| 24 801 | 4 032 | 739 2 | 760 4 | 813 3 | 866 8 | 920 7 | 975 1 |
| 25 044 | 3 993 | 756 9 | 778 0 | 830 8 | 884 3 | 938 0 | 992 3 |
| 25 278 | 3 956 | 776 9 | 799 5 | 856 0 | 913 1 | 970 6 | 1020 6 |
| 25 517 | 3 919 | 795 6 | 816 9 | 870 1 | 923 9 | 978 1 | 1032 8 |
| 25 760 | 3 882 | 815 3 | 836 7 | 890 9 | 944 1 | 998 6 | 1053 4 |
| 26 021 | 3 843 | 836 1 | 875 5 | 911 2 | 965 4 | 1020 0 | 1075 1 |
| 26 274 | 3 806 | 855 8 | 877 3 | 931 1 | 985 5 | 1040 4 | 1095 6 |
| 26 532 | 3 769 | 857 5 | 894 9 | 950 7 | 1004 0 | 1059 8 | 1115 0 |
| 26 802 | 3 731 | 896 0 | 917 4 | 971 0 | 1025 2 | 1079 8 | 1134 9 |
| 27 078 | 3 693 | 910 1 | 931 6 | 984 4 | 1039 8 | 1094 6 | 1149 8 |
| 27 352 | 3 656 | 925 6 | 947 0 | 1000 7 | 1054 9 | 1109 7 | 1164 9 |
| 27 632 | 3 619 | 945 5 | 967 0 | 1020 9 | 1075 4 | 1130 3 | 1185 7 |
| 27 917 | 3 582 | 966 2 | 987 8 | 1043 7 | 1096 6 | 1152 1 | 1207 3 |
| 28 209 | 3 545 | 983 0 | 1004 7 | 1058 9 | 1113 7 | 1169 0 | 1224 7 |
| 28 514 | 3 507 | 1008 4 | 1030 2 | 1084 7 | 1139 8 | 1195 3 | 1251 2 |
| 28 827 | 3 469 | 1029 5 | 1051 4 | 1106 1 | 1161 4 | 1217 1 | 1273 4 |
| 29 137 | 3 432 | 1050 8 | 1072 7 | 1127 7 | 1183 2 | 1239 2 | 1295 6 |
| 29 455 | 3 395 | 1066 8 | 1088 7 | 1143 5 | 1198 9 | 1254 7 | 1311 0 |
| 29 771 | 3 359 | 1087 6 | 1109 7 | 1164 6 | 1220 2 | 1276 3 | 1332 8 |
| 30 120 | 3 320 | 1110 1 | 1132 2 | 1187 4 | 1243 2 | 1300 0 | 1356 2 |
| 30 460 | 3 283 | 1131 9 | 1154 1 | 1209 5 | 1265 5 | 1322 0 | 1378 9 |
| 34 305 | 2 915 | 1356 7 | 1379 8 | 1427 5 | 1495 8 | 1554 5 | 1613 8 |
| 39 246 | 2 549 | 1591 1 | 1614 8 | 1674 3 | 1734 3 | 1794 9 | 1855 9 |
| 45 766 | 2 185 | 1842 3 | 1867 0 | 1928 7 | 1991 0 | 2053 9 | 2117 3 |
| 54 824 | 1 824 | 2072 7 | 2098 0 | 2161 2 | 2225 1 | 2289 5 | 2354 4 |
| 70 126 | 1 466 | 2198 9 | 2225 2 | 2291 0 | 2357 6 | 2424 6 | 2402 2 |

IV

the Slurry Moisture is Retained Constant at 30 per Cent., but the Entering Air by the Outgoing Clinker

per lb. entering at 60° F. Clinkering temperature assumed to be 2,500° F.

| Entering Air at 600° F. Corresponding Exit Temp. | Entering Air at 700° F. Corresponding Exit Temp. | Entering Air at 800° F. Corresponding Exit Temp. | Entering Air at 900° F. Corresponding Exit Temp. | Entering Air at 1,000° F. Corresponding Exit Temp. | Entering Air at Maximum Temp. to which it can be Preheated by Outcoming Clinker. | | Entering Air at 2,500° F. Corresponding Exit Temp. |
|--|--|--|--|--|--|-----------------------|--|
| | | | | | Corresponding Exit Temp. of Gas. | Maximum Temp. of Air. | |
| IX. | X. | XI. | XII. | XIII. | XIV. | XV. | XVI. |
| ° F. | ° F. | ° F. | ° F. | ° F. | ° F. | ° F. | ° F. |
| ... | ... | ... | ... | ... | ... | ... | 188.4 |
| ... | ... | ... | ... | ... | 30.8 | 1944 | 282.9 |
| ... | ... | ... | ... | ... | (imposs.) | | |
| ... | ... | ... | ... | ... | 129.3 | 1853 | 401.1 |
| ... | ... | ... | ... | ... | 231.1 | 1764 | 570.0 |
| ... | ... | ... | ... | 25.6 | 338.7 | 1674 | 744.6 |
| ... | ... | ... | ... | (imposs.) | | | |
| ... | 35.2 | 81.0 | 127.0 | 173.5 | 452.2 | 1584 | 915.2 |
| ... | (imposs.) | | | | | | |
| 139.5 | 186.3 | 233.5 | 281.0 | 329.1 | 571.7 | 1494 | 1095.0 |
| 298.7 | 347.0 | 395.9 | 444.9 | 494.6 | 698.4 | 1403 | 1286.4 |
| 465.6 | 515.5 | 565.6 | 618.6 | 662.2 | 832.3 | 1315 | 1486.9 |
| 646.9 | 698.8 | 751.1 | 803.7 | 856.9 | 992.6 | 1226 | 1705.7 |
| 852.8 | 906.2 | 960.5 | 1014.9 | 1070.1 | 1146.1 | 1137 | 1948.1 |
| 1030.1 | 1085.4 | 1141.3 | 1197.4 | 1254.2 | 1280.9 | 1047 | 2159.2 |
| 1047.2 | 1102.4 | 1158.1 | 1214.2 | 1270.8 | 1294.2 | 1041 | 2174.5 |
| 1067.3 | 1126.3 | 1179.9 | 1235.8 | 1290.5 | 1311.6 | 1031 | 2192.6 |
| 1088.1 | 1143.7 | 1199.9 | 1256.3 | 1313.4 | 1327.1 | 1024 | 2222.8 |
| 1109.0 | 1164.7 | 1221.2 | 1277.8 | 1335.2 | 1343.9 | 1015 | 2249.3 |
| 1130.9 | 1186.8 | 1243.4 | 1300.3 | 1357.8 | 1361.4 | 1006 | 2275.1 |
| 1151.6 | 1207.8 | 1264.6 | 1321.6 | 1379.4 | 1377.6 | 997 | 2300.0 |
| 1170.9 | 1227.0 | 1283.8 | 1337.8 | 1398.5 | 1391.6 | 988 | 2318.2 |
| 1190.6 | 1246.5 | 1303.1 | 1359.9 | 1417.5 | 1405.5 | 979 | 2334.3 |
| 1205.7 | 1261.9 | 1317.4 | 1375.8 | 1433.5 | 1416.1 | 970 | 2353.6 |
| 1220.7 | 1276.7 | 1333.4 | 1390.3 | 1448.0 | 1426.2 | 962 | 2366.4 |
| 1241.8 | 1297.9 | 1354.9 | 1412.0 | 1469.8 | 1442.6 | 953 | 2391.5 |
| 1263.5 | 1320.0 | 1377.1 | 1434.4 | 1492.5 | 1460.0 | 944 | 2417.7 |
| 1281.2 | 1337.7 | 1395.0 | 1452.5 | 1510.7 | 1472.9 | 935 | 2438.5 |
| 1307.9 | 1364.8 | 1422.4 | 1480.1 | 1538.5 | 1495.2 | 926 | 2470.4 |
| 1330.2 | 1387.3 | 1445.1 | 1503.1 | 1561.9 | 1509.5 | 917 | 2497.9 |
| 1352.7 | 1410.0 | 1468.1 | 1526.3 | 1585.2 | 1542.1 | 910 | 2523.8 |
| 1368.0 | 1425.2 | 1483.1 | 1541.2 | 1600.0 | 1540.5 | 899 | 2537.4 |
| 1390.0 | 1447.4 | 1505.5 | 1563.8 | 1622.8 | 1558.6 | 891 | 2564.0 |
| 1413.7 | 1471.3 | 1529.6 | 1588.2 | 1647.5 | 1577.0 | 881 | 2592.2 |
| 1436.6 | 1494.4 | 1553.0 | 1611.8 | 1671.2 | 1595.2 | 872 | 2619.5 |
| 1473.7 | 1733.9 | 1794.8 | 1856.0 | 1917.9 | 1783.4 | 781 | 2904.6 |
| 1917.7 | 1979.7 | 2042.4 | 2105.5 | 2169.2 | 1976.0 | 694 | 3185.8 |
| 2181.5 | 2245.9 | 2311.0 | 2376.5 | 2442.7 | 2184.8 | 605 | 3498.4 |
| 2420.1 | 2486.0 | 2552.8 | 2619.8 | 2687.6 | 2364.8 | 516 | 3768.7 |
| 2560.6 | 2629.3 | 2698.9 | 2768.6 | 2839.3 | 2443.4 | 428 | 3965.1 |

TABLE

Showing the Exit Temperatures of the Gases Issuing from a Rotary Kiln when
 is Preheated to Various Degrees
 Air Supply—10 478 lbs per 1 lb of standard coal of 12,600 B Th U's

| Clinker Production | | Entering Air at 60° F Corre sponding Exit Temp | Entering Air at 100° F Correspond ing Exit Temp | Entering Air at 200° F Correspond ing Exit Temp | Entering Air at 300° F Correspond ing Exit Temp | Entering Air at 400° F Correspond ing Exit Temp | Entering Air at 500° F Correspond ing Exit Temp |
|--|---|--|--|--|--|--|--|
| Tons of Standard Coal per 100 Tons of Clinker | Lbs of Clinker per 1 Lb of Standard Coal | | | | | | |
| I | II | III | IV | V | VI | VII | VIII |
| | | °F | °F | °F | °F | °F | °F |
| 11 893 | 8 408 | | | | | | |
| 12 525 | 7 984 | | | | | | |
| 13 210 | 7 570 | | | | | | |
| 13 957 | 7 165 | Limits of Waste Heat Boiler | | | | | 18 5 (imposs) |
| 14 791 | 6 761 | | | | 54 9 | 114 5 | 165 4 |
| 15 726 | 6 359 | 91 0 | 111 3 | 162 3 | 213 8 | 265 6 | 318 0 |
| 16 770 | 5 963 | 244 0 | 264 9 | 317 2 | 370 0 | 423 3 | 477 0 |
| 17 960 | 5 568 | 404 9 | 426 4 | 480 1 | 534 5 | 589 7 | 644 5 |
| 19 301 | 5 181 | 571 9 | 594 0 | 649 2 | 705 1 | 761 4 | 818 1 |
| 20 855 | 4 795 | 747 4 | 770 1 | 826 9 | 884 4 | 942 2 | 1000 5 |
| 22 660 | 4 413 | 931 4 | 954 8 | 1013 7 | 1072 3 | 1131 9 | 1191 9 |
| 24 801 | 4 032 | 1120 2 | 1144 1 | 1203 9 | 1264 4 | 1325 3 | 1386 7 |
| 25 044 | 3 993 | 1135 2 | 1159 0 | 1218 7 | 1278 9 | 1340 1 | 1403 9 |
| 25 278 | 3 956 | 1156 1 | 1180 1 | 1240 0 | 1300 5 | 1361 6 | 1423 1 |
| 25 517 | 3 919 | 1174 5 | 1198 5 | 1258 5 | 1319 2 | 1380 7 | 1441 8 |
| 25 760 | 3 882 | 1194 2 | 1218 2 | 1278 4 | 1339 2 | 1400 5 | 1462 2 |
| 26 021 | 3 843 | 1214 7 | 1238 8 | 1299 7 | 1360 1 | 1421 6 | 1483 6 |
| 26 274 | 3 806 | 1234 3 | 1258 5 | 1319 0 | 1380 1 | 1441 8 | 1503 9 |
| 26 532 | 3 769 | 1247 9 | 1272 0 | 1332 3 | 1393 2 | 1454 6 | 1516 6 |
| 26 802 | 3 731 | 1260 7 | 1284 7 | 1344 9 | 1405 6 | 1466 9 | 1528 6 |
| 27 078 | 3 693 | 1283 2 | 1307 3 | 1367 6 | 1428 5 | 1489 9 | 1551 9 |
| 27 352 | 3 656 | 1295 4 | 1319 6 | 1379 7 | 1440 5 | 1502 6 | 1563 3 |
| 27 632 | 3 619 | 1315 6 | 1339 7 | 1400 0 | 1460 9 | 1522 3 | 1584 7 |
| 27 917 | 3 582 | 1335 8 | 1360 0 | 1420 4 | 1481 5 | 1543 6 | 1605 3 |
| 28 209 | 3 545 | 1353 2 | 1377 4 | 1438 1 | 1499 3 | 1561 0 | 1623 7 |
| 28 514 | 3 507 | 1377 0 | 1401 3 | 146 1 | 1523 6 | 1585 6 | 1648 9 |
| 28 827 | 3 469 | 1398 2 | 1422 6 | 1483 6 | 1545 3 | 1607 4 | 1670 1 |
| 29 137 | 3 432 | 1418 5 | 1443 0 | 1504 7 | 1566 0 | 1628 0 | 1691 3 |
| 29 455 | 3 395 | 1431 7 | 1456 1 | 1517 1 | 1578 7 | 1640 9 | 1703 5 |
| 29 771 | 3 359 | 1452 3 | 1476 8 | 1538 0 | 1599 8 | 1662 1 | 1725 0 |
| 30 120 | 3 320 | 1474 0 | 1498 5 | 1560 0 | 1622 0 | 1685 0 | 1747 5 |
| 30 460 | 3 283 | 1495 4 | 1520 0 | 1581 6 | 1643 8 | 1706 5 | 1769 8 |
| 34 305 | 2 915 | 1712 0 | 1737 4 | 1800 9 | 1865 1 | 1929 8 | 1995 1 |
| 39 246 | 2 549 | 1929 0 | 1954 9 | 2019 7 | 2085 2 | 2151 1 | 2217 7 |
| 45 766 | 2 185 | 2161 0 | 2187 6 | 2254 3 | 2321 6 | 2389 5 | 2458 0 |
| 54 824 | 1 824 | 2363 1 | 2390 1 | 2457 8 | 2526 1 | 2595 0 | 2664 5 |
| 70 176 | 1 466 | 2622 7 | 2650 5 | 2720 2 | 2790 6 | 2861 5 | 2933 0 |

V

the Slurry Moisture is Retained Constant at 20 per Cent., but the Entering Air by the Outgoing Clinker

per lb. entering at 60° F. Clinkering temperature assumed to be 2,500° F.

| Entering Air at 600° F. Corresponding Exit Temp. | Entering Air at 700° F. Corresponding Exit Temp. | Entering Air at 800° F. Corresponding Exit Temp. | Entering Air at 900° F. Corresponding Exit Temp. | Entering Air at 1,000° F. Corresponding Exit Temp. | Entering Air at Maximum Temp. to which it can be Preheated by Outcoming Clinker. | | Entering Air at 2,500° F. Corresponding Exit Temp. |
|--|--|--|--|--|--|-----------------------|--|
| | | | | | Corresponding Exit Temp. of Gas. | Maximum Temp. of Air. | |
| IX. | X. | XI. | XII. | XIII. | XIV. | XV. | XVI. |
| ° F. | ° F. | ° F. | ° F. | ° F. | ° F. | ° F. | ° F. |
| ... | ... | ... | ... | ... | 338.0 | 2039 | 598.0 |
| ... | ... | ... | ... | ... | 457.0 | 1944 | 773.1 |
| ... | ... | 27.3 (imposs.) | 77.3 | 127.9 | 574.0 | 1853 | 940.0 |
| 68.8 | 119.3 | 170.4 | 221.6 | 272.3 | 687.8 | 1764 | 1104.8 |
| 217.0 | 268.8 | 321.2 | 373.8 | 427.0 | 796.5 | 1674 | 1274.5 |
| 370.9 | 424.0 | 477.8 | 531.8 | 586.5 | 913.8 | 1584 | 1457.6 |
| 531.4 | 586.0 | 641.3 | 696.7 | 752.7 | 1036.1 | 1494 | 1647.0 |
| 700.3 | 756.4 | 813.2 | 870.2 | 927.9 | 1164.5 | 1403 | 1847.3 |
| 875.5 | 932.2 | 991.5 | 1050.1 | 1109.4 | 1298.8 | 1315 | 2054.3 |
| 1059.8 | 1118.9 | 1179.0 | 1239.2 | 1300.1 | 1455.4 | 1226 | 2271.7 |
| 1252.9 | 1313.8 | 1375.5 | 1433.4 | 1500.2 | 1586.9 | 1137 | 2500.4 |
| 1448.9 | 1511.3 | 1574.5 | 1637.9 | 1702.1 | 1732.3 | 1047 | 2725.2 |
| 1462.9 | 1525.1 | 1588.1 | 1651.3 | 1715.3 | 1741.7 | 1041 | 2735.2 |
| 1485.4 | 1544.7 | 1611.2 | 1674.7 | 1739.1 | 1760.4 | 1031 | 2764.0 |
| 1504.2 | 1562.0 | 1630.2 | 1693.7 | 1758.1 | 1773.5 | 1024 | 2784.2 |
| 1524.8 | 1580.6 | 1651.1 | 1714.9 | 1779.4 | 1789.0 | 1015 | 2808.2 |
| 1546.3 | 1599.0 | 1673.0 | 1737.0 | 1801.7 | 1805.7 | 1006 | 2833.7 |
| 1566.8 | 1617.0 | 1693.9 | 1758.0 | 1822.9 | 1820.9 | 997 | 2858.0 |
| 1579.3 | 1635.4 | 1705.8 | 1769.8 | 1834.4 | 1826.7 | 988 | 2865.7 |
| 1591.1 | 1653.9 | 1717.4 | 1781.1 | 1845.6 | 1832.2 | 979 | 2874.0 |
| 1614.6 | 1674.0 | 1741.1 | 1805.1 | 1869.7 | 1850.3 | 970 | 2901.0 |
| 1625.9 | 1688.6 | 1752.1 | 1815.8 | 1880.3 | 1855.8 | 962 | 2908.6 |
| 1646.8 | 1709.7 | 1773.4 | 1837.3 | 1901.8 | 1871.4 | 953 | 2932.6 |
| 1668.1 | 1731.2 | 1795.1 | 1859.2 | 1924.1 | 1887.7 | 944 | 2957.9 |
| 1686.3 | 1749.5 | 1813.5 | 1877.7 | 1942.7 | 1900.6 | 935 | 2979.5 |
| 1711.3 | 1774.8 | 1839.0 | 1903.5 | 1968.8 | 1920.3 | 926 | 3004.1 |
| 1736.0 | 1797.2 | 1861.7 | 1926.3 | 1991.8 | 1937.5 | 917 | 3035.6 |
| 1754.9 | 1818.7 | 1883.4 | 1948.3 | 2014.0 | 1955.0 | 910 | 3051.3 |
| 1767.0 | 1830.6 | 1895.0 | 1959.8 | 2025.2 | 1959.0 | 899 | 3068.6 |
| 1788.7 | 1852.5 | 1917.2 | 1982.0 | 2047.7 | 1976.2 | 891 | 3094.6 |
| 1811.3 | 1875.4 | 1940.2 | 2005.3 | 2071.1 | 1992.8 | 881 | 3121.0 |
| 1833.8 | 1898.0 | 1963.1 | 2028.3 | 2094.3 | 2010.0 | 872 | 3147.5 |
| 2061.1 | 2127.4 | 2194.5 | 2261.8 | 2330.0 | 2181.8 | 781 | 3416.5 |
| 2265.1 | 2352.6 | 2421.1 | 2489.7 | 2559.3 | 2348.6 | 694 | 3667.4 |
| 2527.3 | 2596.8 | 2667.2 | 2737.8 | 2809.3 | 2530.8 | 605 | 3949.0 |
| 2734.9 | 2805.3 | 2876.8 | 2948.4 | 3021.1 | 2675.7 | 516 | 4178.1 |
| 3005.4 | 3078.1 | 3151.7 | 3225.5 | 3299.9 | 2881.4 | 428 | 4491.4 |

TABLE

Showing the Exit Temperatures of the Gases Issuing from a Rotary Kiln when
is Preheated to Various Degrees

Air Supply—10 478 lbs per 1 lb of standard coal of 12,600 B Th U's

| Clinker Production | | Entering Air at 60° F Corre- sponding Exit Temp | Entering Air at 100° F Correspond- ing Exit Temp | Entering Air at 200° F Correspond- ing Exit Temp | Entering Air at 300° F Correspond- ing Exit Temp | Entering Air at 400° F Correspond- ing Exit Temp | Entering Air at 500° F Correspond- ing Exit Temp |
|--|---|---|---|---|---|---|---|
| Tons of Standard Coal per 100 Tons of Clinker | Lbs of Clinker per 1 Lb of Standard Coal | III | IV | V | VI | VII | VIII |
| I | II | III | IV | V | VI | VII | VIII |
| 11 893 | 8 498 | °F Limits of Waste Heat Boiler | | | | °F | °F |
| 12 525 | 7 984 | | | | | 40 1 (imposs) | 10 2 (imposs) 95 9 |
| 13 210 | 7 570 | | | 74 3 | 130 4 | 187 3 | 243 9 |
| 13 957 | 7 165 | 142 1 | 164 7 | 221 3 | 278 5 | 336 1 | 394 1 |
| 14 791 | 6 761 | 203 4 | 316 5 | 374 2 | 432 6 | 491 4 | 550 7 |
| 15 726 | 6 359 | 449 9 | 473 5 | 532 4 | 592 0 | 652 0 | 712 6 |
| 16 770 | 5 963 | 610 4 | 634 4 | 694 6 | 755 4 | 816 7 | 870 5 |
| 17 960 | 5 568 | 777 6 | 802 2 | 863 6 | 925 7 | 988 0 | 1051 3 |
| 19 301 | 5 181 | 948 7 | 973 8 | 1036 6 | 1100 0 | 1164 0 | 1228 5 |
| 20 855 | 4 795 | 1126 3 | 1151 9 | 1216 1 | 1280 9 | 1346 1 | 1412 1 |
| 22 660 | 4 413 | 1310 0 | 1336 2 | 1401 7 | 1468 0 | 1534 0 | 1602 1 |
| 24 801 | 4 032 | 1492 7 | 1519 3 | 1585 8 | 1652 7 | 1721 4 | 1789 2 |
| 25 044 | 3 903 | 1504 4 | 1530 9 | 1597 2 | 1664 1 | 1732 1 | 1799 6 |
| 25 278 | 3 956 | 1525 6 | 1552 1 | 1618 7 | 1686 0 | 1753 8 | 1822 1 |
| 25 517 | 3 919 | 1542 2 | 1568 8 | 1635 3 | 1702 6 | 1770 8 | 1838 7 |
| 25 760 | 3 882 | 1561 5 | 1588 0 | 1654 3 | 1721 3 | 1788 9 | 1857 0 |
| 26 021 | 3 843 | 1581 5 | 1608 2 | 1675 1 | 1742 6 | 1810 7 | 1879 4 |
| 26 274 | 3 806 | 1600 0 | 1627 6 | 1694 6 | 1762 3 | 1830 5 | 1899 3 |
| 26 537 | 3 769 | 1610 2 | 1636 8 | 1703 5 | 1770 9 | 1838 8 | 1907 3 |
| 26 802 | 3 731 | 1620 0 | 1646 5 | 1713 0 | 1780 0 | 1847 7 | 1915 9 |
| 27 078 | 3 693 | 1640 1 | 1666 7 | 1733 2 | 1800 4 | 1868 2 | 1936 5 |
| 27 352 | 3 650 | 1650 7 | 1677 2 | 1743 5 | 1810 4 | 1878 0 | 1946 0 |
| 27 637 | 3 613 | 1670 0 | 1696 5 | 1762 9 | 1830 1 | 1897 7 | 1965 8 |
| 27 917 | 3 580 | 1689 5 | 1716 1 | 1782 7 | 1848 9 | 1918 3 | 1986 1 |
| 28 209 | 3 545 | 1707 0 | 1733 7 | 1800 4 | 1867 8 | 1935 7 | 2004 2 |
| 28 514 | 3 507 | 1729 0 | 1755 7 | 1822 6 | 1890 2 | 1958 3 | 2027 0 |
| 28 877 | 3 469 | 1749 0 | 1775 8 | 1842 8 | 1910 5 | 1978 8 | 2047 6 |
| 29 137 | 3 432 | 1769 0 | 1795 8 | 1863 0 | 1930 9 | 1999 4 | 2068 4 |
| 29 455 | 3 395 | 1778 0 | 1804 7 | 1871 6 | 1939 2 | 2007 4 | 2086 1 |
| 29 771 | 3 359 | 1797 6 | 1824 4 | 1891 5 | 1959 2 | 2027 5 | 2096 4 |
| 30 120 | 3 320 | 1818 6 | 1845 5 | 1912 7 | 1980 9 | 2049 6 | 2118 0 |
| 30 460 | 3 283 | 1838 2 | 1865 1 | 1937 5 | 2000 5 | 2069 1 | 2138 3 |
| 34 305 | 2 915 | 2042 7 | 2070 3 | 2139 2 | 2208 9 | 2279 1 | 2349 9 |
| 39 246 | 2 549 | 2237 9 | 2265 7 | 2335 4 | 2405 8 | 2476 8 | 2548 3 |
| 45 766 | 2 185 | 2446 6 | 2475 0 | 2546 1 | 2617 9 | 2690 3 | 2763 3 |
| 54 824 | 1 824 | 2619 3 | 2647 9 | 2719 4 | 2791 8 | 2864 6 | 2938 1 |
| 70 126 | 1 466 | 2844 7 | 2873 9 | 2946 8 | 3020 5 | 3094 8 | 3169 7 |

VI

the Slurry Moisture is Retained Constant at 10 per Cent., but the Entering Air by the Outgoing Clinker

per lb. entering at 60° F. Clinkering temperature assumed to be 2,500° F.

| Entering Air at 600° F. Corresponding Exit Temp. | Entering Air at 700° F. Corresponding Exit Temp. | Entering Air at 800° F. Corresponding Exit Temp. | Entering Air at 900° F. Corresponding Exit Temp. | Entering Air at 1,000° F. Corresponding Exit Temp. | Entering Air at Maximum Temp.° to which it can be Preheated by Outcoming Clinker. | | Entering Air at 2,500° F. Corresponding Exit Temp. |
|--|--|--|--|--|---|-----------------------|--|
| | | | | | Corresponding Exit Temp. of Gas. | Maximum Temp. of Air. | |
| IX. | X. | XI. | XII. | XIII. | XIV. | XV. | XVI. |
| ° F. | ° F. | ° F. | ° F. | ° F. | ° F. | ° F. | ° F. |
| 73.6 | 129.2 | 165.5 | 242.0 | 299.2 | 917.7 | 2039 | 1210.7 |
| 152.7 | 211.5 | 266.6 | 324.2 | 382.6 | 955.7 | 1944 | 1312.7 |
| 301.5 | 359.4 | 418.0 | 476.8 | 536.3 | 1063.4 | 1853 | 1485.1 |
| 453.1 | 512.1 | 571.9 | 631.8 | 692.5 | 1175.1 | 1764 | 1660.1 |
| 610.7 | 670.9 | 731.9 | 793.1 | 855.0 | 1284.9 | 1674 | 1842.4 |
| 773.9 | 835.3 | 897.6 | 960.0 | 1023.4 | 1402.2 | 1584 | 2031.9 |
| 941.1 | 1003.9 | 1067.5 | 1130.2 | 1195.7 | 1521.8 | 1494 | 2224.9 |
| 1115.2 | 1179.5 | 1244.5 | 1311.7 | 1375.6 | 1646.1 | 1403 | 2426.9 |
| 1293.7 | 1359.3 | 1425.5 | 1492.1 | 1559.5 | 1774.7 | 1315 | 2633.4 |
| 1478.8 | 1545.7 | 1613.5 | 1681.5 | 1750.3 | 1902.7 | 1226 | 2847.4 |
| 1670.4 | 1738.6 | 1808.4 | 1897.3 | 1947.6 | 2044.8 | 1137 | 3069.0 |
| 1858.3 | 1927.8 | 1998.1 | 2068.7 | 2140.0 | 2173.6 | 1047 | 3278.1 |
| 1868.5 | 1937.6 | 2007.6 | 2077.8 | 2148.9 | 2178.2 | 1041 | 3294.0 |
| 1891.3 | 1960.7 | 2031.0 | 2101.5 | 2172.0 | 2196.6 | 1031 | 3310.9 |
| 1907.9 | 1977.3 | 2047.6 | 2118.1 | 2189.5 | 2206.6 | 1024 | 3327.6 |
| 1925.9 | 1995.1 | 2065.1 | 2135.5 | 2206.6 | 2217.3 | 1015 | 3344.6 |
| 1948.8 | 2018.6 | 2089.2 | 2160.0 | 2231.7 | 2226.2 | 1006 | 3362.5 |
| 1968.9 | 2038.9 | 2109.6 | 2180.6 | 2252.4 | 2240.2 | 997 | 3379.2 |
| 1976.6 | 2046.2 | 2116.6 | 2187.2 | 2258.8 | 2250.2 | 988 | 3397.1 |
| 1984.9 | 2054.2 | 2124.4 | 2194.8 | 2266.0 | 2251.1 | 979 | 3414.3 |
| 2005.7 | 2075.1 | 2145.4 | 2215.9 | 2287.2 | 2265.8 | 970 | 3430.1 |
| 2014.9 | 2084.1 | 2154.1 | 2224.3 | 2295.4 | 2268.5 | 962 | 3446.0 |
| 2035.0 | 2104.2 | 2174.3 | 2244.8 | 2315.9 | 2282.4 | 953 | 3462.0 |
| 2055.3 | 2124.8 | 2195.1 | 2265.7 | 2337.0 | 2297.1 | 944 | 3475.6 |
| 2073.5 | 2143.2 | 2213.6 | 2284.3 | 2355.8 | 2309.4 | 935 | 3496.5 |
| 2096.5 | 2166.2 | 2236.9 | 2307.7 | 2379.5 | 2326.4 | 926 | 3523.0 |
| 2117.3 | 2187.3 | 2258.0 | 2329.1 | 2401.0 | 2341.3 | 917 | 3547.2 |
| 2138.2 | 2208.3 | 2279.4 | 2350.4 | 2422.6 | 2357.7 | 910 | 3572.0 |
| 2145.6 | 2215.4 | 2286.0 | 2357.0 | 2428.7 | 2356.1 | 899 | 3572.8 |
| 2166.1 | 2236.1 | 2306.8 | 2377.9 | 2449.8 | 2371.6 | 891 | 3596.6 |
| 2187.9 | 2258.1 | 2329.1 | 2400.2 | 2472.3 | 2386.6 | 881 | 3621.6 |
| 2208.3 | 2278.6 | 2349.7 | 2421.1 | 2493.3 | 2401.0 | 872 | 3645.2 |
| 2421.5 | 2493.5 | 2566.3 | 2642.0 | 2713.4 | 2552.6 | 781 | 3892.6 |
| 2620.7 | 2693.4 | 2767.0 | 2840.9 | 2915.6 | 2689.1 | 694 | 4107.1 |
| 2837.1 | 2911.3 | 2986.3 | 3061.7 | 3137.9 | 2840.9 | 605 | 4353.4 |
| 3012.6 | 3087.1 | 3162.8 | 3238.5 | 3315.3 | 2950.0 | 516 | 4538.8 |
| 3245.5 | 3321.7 | 3398.7 | 3476.0 | 3554.3 | 3115.7 | 428 | 4801.8 |

TABLE

Showing the Exit Temperatures of the Gases Issuing from a Rotary Kiln when
is Preheated to Various Degrees

Air Supply—10 478 lbs. per 1 lb of standard coal of 12,600 B Th U's

| Clinker Production | | Entering Air at 60° F Corre- sponding Exit Temp | Entering Air at 100° F Correspond- ing Exit Temp | Entering Air at 200° F Correspond- ing Exit Temp | Entering Air at 300° F Correspond- ing Exit Temp | Entering Air at 400° F Correspond- ing Exit Temp | Entering Air at 500° F Correspond- ing Exit Temp |
|--------------------|-------|---|---|---|---|---|---|
| I | II | III | IV | V | VI | VII | VIII |
| | | °F | °F | °F | °F | °F | °F |
| 11 893 | 8 408 | 129 2 | 154 3 | 217 0 | 280 3 | 344 7 | 408 6 |
| 12 525 | 7 984 | 202 2 | 227 6 | 291 3 | 355 6 | 420 4 | 485 7 |
| 13 210 | 7 570 | 355 2 | 381 0 | 445 5 | 510 8 | 577 1 | 642 9 |
| 13 957 | 7 165 | 509 3 | 535 5 | 601 0 | 667 1 | 733 8 | 801 1 |
| 14 791 | 6 761 | 667 6 | 694 1 | 760 6 | 927 7 | 895 4 | 963 6 |
| 15 726 | 6 359 | 829 5 | 856 4 | 923 8 | 992 0 | 1060 6 | 1129 8 |
| 16 770 | 5 963 | 993 6 | 1020 9 | 1089 3 | 1158 5 | 1228 1 | 1298 4 |
| 17 960 | 5 568 | 1162 4 | 1190 1 | 1259 6 | 1329 7 | 1401 0 | 1471 7 |
| 19 301 | 5 181 | 1332 9 | 1361 1 | 1431 5 | 1502 7 | 1574 5 | 1646 8 |
| 20 855 | 4 795 | 1507 7 | 1536 3 | 1601 8 | 1680 1 | 1753 5 | 1826 4 |
| 22 660 | 4 413 | 1686 3 | 1715 0 | 1787 6 | 1860 9 | 1934 9 | 2009 4 |
| 24 801 | 4 032 | 1856 9 | 1886 1 | 1959 2 | 2033 1 | 2108 2 | 2182 7 |
| 25 044 | 3 993 | 1864 2 | 1893 2 | 1966 0 | 2039 4 | 2113 4 | 2188 2 |
| 25 278 | 3 956 | 1885 6 | 1914 8 | 1987 8 | 2061 5 | 2135 9 | 2210 8 |
| 25 517 | 3 919 | 1900 1 | 1929 2 | 2002 1 | 2075 7 | 2150 5 | 2224 8 |
| 25 760 | 3 882 | 1918 4 | 1947 6 | 2020 7 | 2094 5 | 2168 9 | 2243 9 |
| 26 021 | 3 843 | 1937 0 | 1966 2 | 2039 4 | 2113 3 | 2187 8 | 2262 9 |
| 26 274 | 3 806 | 1955 4 | 1984 7 | 2057 9 | 2132 0 | 2206 6 | 2281 8 |
| 26 532 | 3 769 | 1960 7 | 1989 8 | 2062 6 | 2136 2 | 2210 4 | 2285 2 |
| 26 802 | 3 731 | 1965 9 | 1994 9 | 2067 3 | 2140 5 | 2244 3 | 2268 7 |
| 27 078 | 3 693 | 1986 5 | 2015 5 | 2088 1 | 2161 4 | 2235 3 | 2309 8 |
| 27 352 | 3 656 | 1992 2 | 2021 1 | 2093 3 | 2166 2 | 2239 7 | 2313 9 |
| 27 632 | 3 619 | 2009 7 | 2038 6 | 2110 9 | 2184 0 | 2257 6 | 2331 8 |
| 27 917 | 3 582 | 2028 8 | 2057 7 | 2130 1 | 2203 3 | 2277 9 | 2351 4 |
| 28 209 | 3 545 | 2045 2 | 2074 2 | 2146 7 | 2219 9 | 2293 8 | 2368 3 |
| 28 514 | 3 507 | 2064 8 | 2093 8 | 2166 5 | 2239 8 | 2313 8 | 2388 3 |
| 28 827 | 3 469 | 2083 4 | 2112 6 | 2185 3 | 2258 8 | 2332 9 | 2407 6 |
| 29 137 | 3 432 | 2101 8 | 2130 9 | 2203 8 | 2277 4 | 2351 6 | 2426 4 |
| 29 455 | 3 395 | 2107 5 | 2136 5 | 2209 0 | 2282 2 | 2356 0 | 2430 5 |
| 29 771 | 3 359 | 2125 8 | 2154 8 | 2227 4 | 2300 8 | 2374 8 | 2449 3 |
| 30 120 | 3 320 | 2144 8 | 2174 9 | 2250 1 | 2326 1 | 2403 3 | 2480 0 |
| 30 460 | 3 283 | 2163 3 | 2192 4 | 2265 2 | 2338 8 | 2413 0 | 2487 8 |
| 34 305 | 2 915 | 2350 4 | 2380 0 | 2454 0 | 2528 7 | 2604 0 | 2680 1 |
| 39 246 | 2 549 | 2520 1 | 2549 7 | 2623 9 | 2698 8 | 2774 3 | 2850 0 |
| 45 766 | 2 185 | 2704 0 | 2734 0 | 2809 0 | 2884 8 | 2961 2 | 3038 2 |
| 54 824 | 1 824 | 2846 2 | 2876 2 | 2951 " | 3027 1 | 3103 5 | 3180 5 |
| 70 126 | 1 466 | 3037 9 | 3068 2 | 3144 0 | 3220 6 | 3297 8 | 3375 7 |

VII

the Slurry Moisture is Retained Constant at 0 per Cent., but the Entering Air by the Outgoing Clinker

per lb. entering at 60° F. Clinkering temperature assumed to be 2,500° F.

| IX. | X. | XI. | XII. | XIII. | Entering Air at Maximum Temp. to which it can be Preheated by Outcoming Clinker. | | XVI. |
|--------|--------|--------|--------|--------|--|-----------------------|--------|
| | | | | | Corresponding Exit Temp. of Gas. | Maximum Temp. of Air. | |
| ° F. | ° F. | ° F. | ° F. | ° F. | ° F. | ° F. | ° F. |
| 473.2 | 539.2 | 605.5 | 675.5 | 739.2 | 1469.4 | 2039 | 1811.7 |
| 551.9 | 618.3 | 685.5 | 752.9 | 821.2 | 1491.9 | 1944 | 1909.5 |
| 709.9 | 777.3 | 845.5 | 913.9 | 983.1 | 1596.1 | 1853 | 2087.2 |
| 869.2 | 937.5 | 1006.7 | 1076.1 | 1146.3 | 1702.3 | 1764 | 2266.3 |
| 1032.6 | 1102.0 | 1172.8 | 1242.5 | 1313.7 | 1807.6 | 1674 | 2450.6 |
| 1199.9 | 1270.2 | 1341.4 | 1412.8 | 1485.2 | 1918.2 | 1584 | 2642.8 |
| 1369.6 | 1440.9 | 1513.2 | 1585.7 | 1659.0 | 2029.6 | 1494 | 2828.6 |
| 1543.9 | 1616.4 | 1689.7 | 1763.3 | 1837.8 | 2146.8 | 1403 | 3020.4 |
| 1720.1 | 1793.6 | 1868.0 | 1942.7 | 2018.3 | 2259.9 | 1315 | 3223.3 |
| 1900.7 | 1975.3 | 2050.9 | 2126.7 | 2203.4 | 2379.5 | 1226 | 3426.5 |
| 2085.1 | 2160.6 | 2237.3 | 2314.3 | 2392.0 | 2499.6 | 1137 | 3634.0 |
| 2258.7 | 2335.0 | 2412.2 | 2489.8 | 2568.2 | 2605.1 | 1047 | 3773.1 |
| 2263.7 | 2339.6 | 2416.4 | 2493.5 | 2571.5 | 2613.7 | 1041 | 3794.9 |
| 2286.7 | 2362.9 | 2439.9 | 2517.3 | 2595.6 | 2621.6 | 1031 | 3813.8 |
| 2300.5 | 2376.6 | 2453.5 | 2530.8 | 2608.9 | 2627.6 | 1024 | 3825.1 |
| 2319.9 | 2396.1 | 2473.2 | 2550.7 | 2628.9 | 2636.8 | 1015 | 3844.4 |
| 2338.9 | 2415.3 | 2492.5 | 2570.0 | 2648.5 | 2645.4 | 1006 | 3863.0 |
| 2358.0 | 2434.4 | 2511.8 | 2589.4 | 2668.0 | 2664.6 | 997 | 3882.0 |
| 2360.9 | 2436.9 | 2513.8 | 2591.0 | 2669.2 | 2663.8 | 988 | 3894.0 |
| 2364.0 | 2439.6 | 2516.1 | 2592.9 | 2670.6 | 2672.4 | 979 | 3903.6 |
| 2385.3 | 2461.0 | 2537.6 | 2614.5 | 2692.4 | 2681.9 | 970 | 3920.4 |
| 2389.0 | 2464.3 | 2540.6 | 2617.0 | 2694.4 | 2690.1 | 962 | 3939.2 |
| 2407.0 | 2482.5 | 2558.8 | 2635.4 | 2712.0 | 2697.5 | 953 | 3946.7 |
| 2426.7 | 2502.3 | 2578.7 | 2655.5 | 2733.1 | 2704.5 | 944 | 3961.6 |
| 2443.7 | 2519.3 | 2595.9 | 2672.7 | 2750.4 | 2711.0 | 935 | 3980.5 |
| 2463.8 | 2539.6 | 2616.3 | 2693.3 | 2771.2 | 2718.3 | 926 | 4002.8 |
| 2483.3 | 2559.2 | 2638.8 | 2713.1 | 2791.1 | 2727.3 | 917 | 4020.3 |
| 2502.1 | 2578.1 | 2655.1 | 2732.2 | 2810.4 | 2736.3 | 910 | 4038.8 |
| 2505.8 | 2581.5 | 2658.0 | 2734.9 | 2812.6 | 2745.7 | 899 | 4052.4 |
| 2524.8 | 2600.6 | 2677.2 | 2754.2 | 2832.0 | 2753.7 | 891 | 4073.8 |
| 2558.2 | 2636.7 | 2713.1 | 2790.9 | 2856.6 | 2760.7 | 881 | 4095.2 |
| 2563.5 | 2639.5 | 2719.4 | 2793.6 | 2871.9 | 2771.9 | 872 | 4117.4 |
| 2757.0 | 2834.2 | 2912.3 | 2993.5 | 3070.2 | 2897.6 | 781 | 4335.5 |
| 2927.6 | 3004.9 | 3083.4 | 3161.9 | 3241.5 | 3000.4 | 694 | 4509.9 |
| 3116.3 | 3194.5 | 3273.8 | 3353.3 | 3433.7 | 3120.3 | 605 | 4716.5 |
| 3258.5 | 3336.8 | 3416.1 | 3495.5 | 3576.0 | 3193.0 | 516 | 4859.2 |
| 3454.5 | 3533.6 | 3613.8 | 3694.0 | 3775.3 | 3319.5 | 428 | 5071.9 |

CHAPTER XXVI

CALCULATION OF THE AMOUNT OF STEAM
PRODUCIBLE PER 100 TONS OF CLINKER IN
THE ROTARY KILN, TOGETHER WITH A
DISCUSSION OF THE WASTE-HEAT BOILERAmount of Steam Produisible by the Exit Gases of a
Cement Rotary Kiln

§ 1. In the preceding Chapter (XXV.) were calculated the exit temperatures of the gases coming from cement rotary kilns using slurries containing different percentages of moisture (0 to 40 per cent.) and the kiln working with different fuel consumptions.

In the present chapter it is proposed to calculate the amount of steam producible by utilising the waste heat contained in these gases.

For practical purposes it is sufficient to consider that the steam is produced in a waste-heat boiler at a pressure of 150 lbs. per sq. in. (by gauge). This corresponds to a temperature of about 366° F. inside the boiler. Hence the steam at this pressure can only be generated by the heat liberated by the exit gases in sinking from their exit temperature T° to 366° F.

Now, the number of B.Th.U.'s absorbed in turning 1 lb. of water at 366° F. contained in the boiler under 150 lbs. pressure (by gauge) into 1 lb. of steam at the same temperature and pressure is equal to the latent heat of evaporation of water under these conditions, being, according to Callendar's Steam Tables, = 863.1 B.Th.U.'s.

Hence, if Q B.Th.U.'s be the quantity of heat available from a given quantity of gas in sinking from T° to 366° F., then the amount of steam producible therefrom is

$$\frac{Q}{863.1} = Q \times 0.001159 \text{ lb.} \quad . \quad . \quad . \quad . \quad (1)$$

This is the mode of calculation adopted in the following pages.

§ 2. The weight of steam thus calculated is obviously the theoretical maximum. The practical amount will be less than this on account of the varying efficiencies of the boilers used in generating steam. Hence a deduction (depending upon the efficiency of the boiler used) must be made to obtain practical figures.

For some purposes (*e.g.*, when the steam is to be superheated) the practical engineer may prefer to know the value of the available B.Th.U.'s per lb. of clinker which can be utilised for steam generation above 366° F. This figure can easily be calculated from the weights of steam given in the table by multiplying the weight of steam given by 863.1, the latent heat of steam at 366° F. and 150 lbs. pressure (by gauge), *e.g.*, if to 1 lb. of clinker we can obtain 1.2 lbs. of steam at 366° F. and 150 lbs. pressure, then the number of B.Th.U.'s available for steam generation at 150 lbs. is

$$1.2 \times 863.1 = 1,036 \text{ B.Th.U.'s.}$$

If the steam is to be superheated to T° and its specific heat is S , we have, if W be the weight of steam

$$1036 = W(863 + S(T - 366)),$$

or

$$W = \frac{1036}{863 + S(T - 366)} \quad (1)$$

§ 3 Details of Calculating the Amount of Heat Q Contained in the Exit Gases per 1 Lb of Clinker Produced by the Kiln—For practical purposes it is convenient to calculate the lbs of steam producible per 1 lb of clinker produced in the kiln. The method is best illustrated by working through a particular case in detail, in which use is made of the data calculated in the preceding chapters.

Take the following case. Suppose that the rotary kiln is consuming 30460 tons of standard coal (12,600 B Th U's per lb) per 100 lbs clinker produced. This amounts to 3283 lbs of clinker per 1 lb of standard coal burnt.

Suppose that the slurry contains 40 per cent of moisture and the air meets the coal dust preheated to 400°F , then the quantities associated with these numbers (calculated on 3283 lbs of clinker) are

| | |
|---|-----------|
| Furnace gases, 11278 lbs | |
| CO_2 evolved from slurry, 1720 lbs | |
| H_2O evolved (a) from kaolin | 06844 lb |
| (b) „ silica | 00355 „ |
| (c) „ slurry | 34150 lbs |
| Total H_2O | 35349 lbs |
| Exit temperature of kiln gases, 914°F | |

The instantaneous specific heat of the furnace gases at 914°F is 02640, and at 366°F is 02471, thus giving a mean specific heat of the furnace gases between 914° and 366°F of 02556 $^\circ\text{F}$.

The maximum temperature drop of the gases between entering the boiler and emerging therefrom is

$$\text{Exit temperature} - 366^\circ\text{F} = 914^\circ - 366^\circ\text{F} = 548^\circ\text{F}.$$

The quantity of heat Q available is now calculated as follows—

(a) The B Th U's evolved when 11278 lbs of furnace gases sink from 914° to 366°F are

$$11278 \times 548 \times 02556 = 1579976 \text{ B Th U's}$$

(b) From our heat tables, the B Th U's liberated when 1720 lbs of CO_2 sink from 914° to 366°F are

$$1720(205633 - 71573) = 230583 \text{ B Th U's}$$

(c) From our heat tables, the B Th U's liberated when 3535 lbs of steam sink from 914° to 366°F are

$$3535(414446 - 156116) = 913197 \text{ B Th U's}$$

Hence the total number of B Th U's liberated when the temperature of exit gases sinks from 914° to 366°F is

$$1579976 + 230583 + 913197 = 2723756 \text{ B Th U's}$$

Consequently, the number of lbs of steam producible will be

$$2723756 \times 0001159 = 3157 \text{ lbs}$$

Hence for every 3283 lbs of clinker produced we can obtain 3157 lbs of steam, or for every 1 lb of clinker we can obtain 0616 lb of steam.

§ 4. This calculation has been repeated for various yields of clinker for kilns having fuel consumptions ranging from 11.893 to 70.126 tons of standard coal per 100 tons of clinker, and fed with slurries containing successively 0, 10, 20, 30, and 40 per cent. of water, and with the entering air preheated to 100°, 400°, 800°, 1,000° F., and to the maximum temperature it could obtain if all the heat in the issuing clinker was imparted to the entering air.

The values of the lbs. of steam producible under these circumstances are set forth in the tables below.

By interpolation from the results contained in these tables it is possible to calculate in a very simple manner the weight of steam producible under almost any other circumstances for any fuel consumption, slurry moisture, and degree of preheating of the entering air.

The general method of interpolation was explained in the preceding chapter on "Exit Temperatures."

§ 5. **Practical Results of the Tables.**—On surveying the tables the following facts become apparent:—

- (a) The more efficient the rotary kiln, the smaller the quantity of steam producible per 1 lb. of clinker.
- (b) Assuming equal fuel consumptions, the amount of steam producible rapidly increases as we decrease the quantity of slurry moisture.
- (c) Even in the most efficient modern rotary kilns using slurry moisture of 40 per cent. a very large amount of high-pressure steam is producible.

Unless a radical change of design is introduced in almost any modern kiln, a waste-heat boiler is a proposition which will repay careful consideration from practical engineers.

If, however, a kiln is designed of such high efficiency as regards clinker production when fed with a slurry containing 40 per cent. moisture it burns only about 21 tons of standard coal, the exit temperature will be so low that it will be impossible to attach a waste-heat boiler to a kiln.

Note.—These tables are calculated from interpolated values of specific heats, and therefore can only be regarded as sufficiently accurate for works engineers.

TABLE

Showing the Number of Lbs of Steam at 150 Lbs Pressure by Gauge per Square Retained Constant at 0 per Cent, but the Entering Air Air Supply—10 478 lbs per 1 lb of standard coal of 12,600 B Th U s

| Clinker Production | | Entering Combustion Air Preheated to 100° F | | Entering Combustion Air Preheated to 400° F | |
|--|--|--|--|--|--|
| Tons of Standard Coal per 100 Tons of Clinker I | Lbs of Clinker per 1 Lb of Standard Coal II | Exit Temperature of Kila Gases III | Number of Lbs Steam Producible per 1 Lb of Clinker IV | Exit Temperature of Kila Gases V | Number of Lbs Steam Producible per 1 Lb of Clinker VI |
| | | ° F | | ° F | |
| 11 893 | 8 408 | | Imposs | 344 7 | Imposs |
| 12 525 | 7 984 | 227 6 | , | 420 4 | 0 0402 |
| 13 210 | 7 570 | 381 0 | 0 01717 | 577 1 | 0 1253 |
| 13 957 | 7 165 | 535 5 | 0 1050 | 733 8 | 0 2306 |
| 14 791 | 6 761 | 694 1 | 0 2137 | 895 4 | 0 3498 |
| 15 726 | 6 359 | 856 4 | 0 3385 | 1060 6 | 0 4871 |
| 16 770 | 5 963 | 1020 9 | 0 4808 | 1228 1 | 0 6416 |
| 17 960 | 5 568 | 1190 1 | 0 6451 | 1401 0 | 0 8211 |
| 19 301 | 5 181 | 1361 1 | 0 8318 | 1574 5 | 1 0233 |
| 20 855 | 4 795 | 1536 3 | 1 0522 | 1753 5 | 1 2595 |
| 22 660 | 4 413 | 1715 0 | 1 310 | 1934 9 | 1 5404 |
| 24 801 | 4 032 | 1886 1 | 1 603 | 2108 2 | 1 855 |
| 25 044 | 3 993 | 1893 2 | 1 623 | 2113 4 | 1 887 |
| 25 278 | 3 956 | 1914 8 | 1 663 | 2135 9 | 1 919 |
| 25 517 | 3 919 | 1929 2 | 1 691 | 2150 1 | 1 957 |
| 25 760 | 3 822 | 1947 6 | 1 726 | 2168 9 | 1 986 |
| 26 021 | 3 843 | 1966 2 | 1 763 | 2187 8 | 2 027 |
| 26 274 | 3 806 | 1984 7 | 1 798 | 2206 6 | 2 068 |
| 26 532 | 3 769 | 1989 8 | 1 822 | 2210 4 | 2 089 |
| 26 802 | 3 731 | 1994 9 | 1 841 | 2214 3 | 2 111 |
| 27 078 | 3 693 | 2015 5 | 1 882 | 2235 3 | 2 155 |
| 27 352 | 3 656 | 2021 1 | 1 905 | 2239 7 | 2 179 |
| 27 632 | 3 619 | 2038 6 | 1 946 | 2257 6 | 2 222 |
| 27 917 | 3 582 | 2057 7 | 1 984 | 2277 9 | 2 266 |
| 28 209 | 3 545 | 2074 2 | 2 025 | 2293 8 | 2 305 |
| 28 514 | 3 507 | 2093 8 | 2 068 | 2313 8 | 2 354 |
| 28 827 | 3 469 | 2112 6 | 2 114 | 2332 9 | 2 400 |
| 29 137 | 3 432 | 2130 9 | 2 155 | 2351 6 | 2 449 |
| 29 455 | 3 395 | 2136 5 | 2 183 | 2356 0 | 2 473 |
| 29 771 | 3 359 | 2154 8 | 2 228 | 2374 8 | 2 524 |
| 30 120 | 3 320 | 2174 9 | 2 273 | 2403 3 | 2 589 |
| 30 460 | 3 283 | 2192 4 | 2 320 | 2413 0 | 2 631 |
| 34 395 | 2 915 | 2380 0 | 2 862 | 2604 0 | 3 211 |
| 39 246 | 2 549 | 2549 7 | 3 515 | 2774 3 | 3 912 |
| 45 766 | 2 185 | 2734 0 | 4 404 | 2961 2 | 4 878 |
| 54 824 | 1 874 | 2876 2 | 5 533 | 3103 5 | 6 100 |
| 70 126 | 1 466 | 3068 2 | 7 348 | 3297 8 | 8 058 |

I

Inch Produccible per 1 Lb. of Clinker Produced when the Slurry Moisture is is Preheated to Various Degrees by the Outgoing Clinker.

per 1 lb. entering at 60° F. Clunkering temperature assumed to be 2,500° F.

| Entering Combustion. Air Preheated to 800° F. | | Entering Combustion. Air Preheated to 1,000° F. | | Entering Air at Maximum Tempera- ture to which it can be Preheated by Outgoing Clinker. | | |
|---|---|--|--|---|---|---|
| Exit Tempera- ture of Kiln Gases. VII. | Number of Lbs. Steam Produceible per 1 Lb. of Clinker. VIII. | Exit Tempera- ture of Kiln Gases. IX. | Number of Lbs. Steam Produceible per 1 Lb. of Clinker. X. | Number of Lbs. Steam Produceible per 1 Lb. of Clinker. XI. | Tempera- ture of Entering Combustion Air. XII. | Exit Tempera- ture of Kiln Gases. XIII. |
| ° F. | | ° F. | | | ° F. | ° F. |
| 605.5 | 0.1325 | 739.2 | 0.2080 | 0.6471 | 2039 | 1469.4 |
| 685.5 | 0.1842 | 821.2 | 0.2586 | 0.6866 | 1944 | 1491.9 |
| 845.5 | 0.2900 | 983.1 | 0.3777 | 0.7849 | 1853 | 1596.1 |
| 1006.7 | 0.4093 | 1146.3 | 0.5035 | 0.8911 | 1764 | 1702.3 |
| 1172.8 | 0.5441 | 1313.7 | 0.6423 | 0.8691 | 1674 | 1807.6 |
| 1341.4 | 0.6950 | 1485.2 | 0.8062 | 1.440 | 1584 | 1918.2 |
| 1513.2 | 0.8690 | 1659.0 | 0.9879 | 1.292 | 1494 | 2029.6 |
| 1689.7 | 1.091 | 1837.8 | 1.196 | 1.470 | 1403 | 2146.8 |
| 1868.0 | 1.292 | 2018.3 | 1.429 | 1.659 | 1315 | 2259.9 |
| 2050.9 | 1.556 | 2203.4 | 1.7095 | 1.745 | 1226 | 2379.5 |
| 2237.3 | 1.862 | 2392.0 | 2.060 | 2.149 | 1137 | 2499.6 |
| 2412.2 | 2.209 | 2568.2 | 2.397 | 2.441 | 1047 | 2605.1 |
| 2416.4 | 2.236 | 2571.5 | 2.420 | 2.4652 | 1041 | 2613.7 |
| 2439.9 | 2.280 | 2595.6 | 2.4699 | 2.5022 | 1031 | 2621.6 |
| 2453.5 | 2.314 | 2608.9 | 2.508 | 2.5288 | 1024 | 2627.6 |
| 2473.2 | 2.357 | 2628.9 | 2.550 | 2.563 | 1015 | 2636.8 |
| 2492.5 | 2.393 | 2648.5 | 2.568 | 2.598 | 1006 | 2645.4 |
| 2511.8 | 2.445 | 2668.0 | 2.6412 | 2.635 | 997 | 2654.6 |
| 2513.8 | 2.467 | 2669.2 | 2.6694 | 2.649 | 988 | 2663.8 |
| 2516.1 | 2.490 | 2670.6 | 2.688 | 2.664 | 979 | 2672.4 |
| 2537.6 | 2.542 | 2692.4 | 2.738 | 2.706 | 970 | 2681.9 |
| 2540.6 | 2.567 | 2694.4 | 2.767 | 2.733 | 962 | 2690.1 |
| 2558.8 | 2.611 | 2712.0 | 2.8125 | 2.766 | 953 | 2697.5 |
| 2578.7 | 2.673 | 2733.1 | 2.8353 | 2.807 | 944 | 2704.5 |
| 2595.9 | 2.707 | 2750.4 | 2.914 | 2.848 | 935 | 2711.0 |
| 2616.3 | 2.755 | 2771.2 | 2.965 | 2.892 | 926 | 2718.3 |
| 2638.8 | 2.820 | 2791.1 | 3.024 | 2.932 | 917 | 2727.3 |
| 2655.1 | 2.865 | 2810.4 | 3.080 | 2.982 | 910 | 2736.3 |
| 2658.0 | 2.891 | 2812.6 | 3.113 | 2.996 | 899 | 2745.7 |
| 2677.2 | 2.946 | 2832.0 | 3.166 | 3.035 | 891 | 2753.7 |
| 2713.1 | 3.024 | 2856.6 | 3.217 | 3.124 | 881 | 2760.7 |
| 2719.4 | 3.061 | 2871.9 | 3.286 | 3.1435 | 872 | 2771.9 |
| 2912.3 | 3.704 | 3070.2 | 3.961 | 3.650 | 781 | 2897.6 |
| 3083.4 | 4.483 | 3241.5 | 4.777 | 4.323 | 694 | 3000.4 |
| 3273.8 | 5.541 | 3433.7 | 6.586 | 5.213 | 605 | 3120.3 |
| 3416.1 | 6.902 | 3576.0 | 7.309 | 6.332 | 516 | 3193.0 |
| 3613.8 | 9.075 | 3775.3 | 9.754 | 8.124 | 428 | 3319.5 |

TABLE

Showing the Number of Lbs of Steam at 150 Lbs Pressure by Gauge per Square Retained Constant at 10 per Cent, but the Entering Air Air Supply—10 478 lbs. per 1 lb of standard coal of 12,600 B Th U's

| Clinker Production | | Entering Combustion Air Preheated to 100° F | | Entering Combustion Air Preheated to 400° F | |
|--|--|--|--|--|--|
| Tons of Standard Coal per 100 Tons of Clinker I | Lbs of Clinker per 1 Lb of Standard Coal II | Exit Temperature of Kiln Gases III | Number of Lbs Steam Producible per 1 Lb of Clinker IV | Exit Temperature of Kiln Gases V | Number of Lbs Steam Producible per 1 Lb of Clinker VI |
| | | ° F | | ° F | |
| 11 893 | 8 408 | | Imposs | Imposs | Imposs |
| 12 525 | 7 984 | | " | " | " |
| 13 210 | 7 570 | | " | " | " |
| 13 957 | 7 165 | | " | 336 1 | " |
| 14 791 | 6 761 | | " | 491 4 | 0 09689 |
| 15 726 | 6 359 | 473 5 | 0 0828 | 652 0 | 0 2216 |
| 16 770 | 5 963 | 634 4 | 0 2165 | 816 7 | 0 3553 |
| 17 960 | 5 568 | 802 2 | 0 3732 | 988 9 | 0 5409 |
| 19 301 | 5 181 | 973 8 | 0 5537 | 1164 0 | 0 7367 |
| 20 855 | 4 795 | 1151 9 | 0 7664 | 1346 2 | 0 9664 |
| 22 660 | 4 413 | 1336 2 | 1 015 | 1534 8 | 1 240 |
| 24 801 | 4 032 | 1519 3 | 1 306 | 1721 4 | 1 548 |
| 25 044 | 3 993 | 1530 9 | 1 329 | 1732 1 | 1 576 |
| 25 278 | 3 956 | 1552 1 | 1 365 | 1753 8 | 1 616 |
| 25 517 | 3 919 | 1568 8 | 1 395 | 1770 8 | 1 648 |
| 25 760 | 3 882 | 1588 0 | 1 430 | 1788 9 | 1 680 |
| 26 021 | 3 843 | 1608 2 | 1 465 | 1810 7 | 1 723 |
| 26 274 | 3 806 | 1627 6 | 1 505 | 1830 5 | 1 764 |
| 26 532 | 3 769 | 1636 8 | 1 525 | 1838 8 | 1 786 |
| 26 802 | 3 731 | 1646 5 | 1 549 | 1847 7 | 1 812 |
| 27 078 | 3 693 | 1666 7 | 1 588 | 1868 2 | 1 852 |
| 27 352 | 3 656 | 1677 2 | 1 614 | 1878 0 | 1 887 |
| 27 632 | 3 619 | 1696 5 | 1 652 | 1897 7 | 1 922 |
| 27 917 | 3 582 | 1716 1 | 1 692 | 1918 3 | 1 966 |
| 28 209 | 3 545 | 1733 7 | 1 732 | 1935 7 | 2 011 |
| 28 514 | 3 507 | 1755 7 | 1 775 | 1958 3 | 2 051 |
| 28 827 | 3 469 | 1775 8 | 1 816 | 1978 8 | 2 102 |
| 29 137 | 3 432 | 1795 8 | 1 861 | 1999 4 | 2 148 |
| 29 455 | 3 395 | 1804 7 | 1 906 | 2007 4 | 2 178 |
| 29 771 | 3 359 | 1824 4 | 1 925 | 2027 5 | 2 223 |
| 30 120 | 3 320 | 1845 5 | 1 981 | 2049 6 | 2 279 |
| 30 460 | 3 283 | 1865 1 | 2 025 | 2069 1 | 2 325 |
| 34 305 | 2 915 | 2070 3 | 2 521 | 2279 1 | 2 907 |
| 39 246 | 2 549 | 2265 7 | 3 215 | 2476 8 | 3 607 |
| 45 766 | 2 185 | 2475 0 | 4 082 | 2690 3 | 4 567 |
| 54 824 | 1 824 | 2647 9 | 5 746 | 2864 6 | 5 786 |
| 70 126 | 1 466 | 2873 9 | 7 039 | 3094 8 | 7 721 |

II

Inch Producing per 1 Lb. of Clinker Produced when the Slurry Moisture is Preheated to Various Degrees by the Outgoing Clinker

per lb. entering at 60° F. Clinkering temperature assumed to be 2,500° F.

| Entering Combustion. Air Preheated to 800° F. | | Entering Combustion. Air Preheated to 1,000° F. | | Entering Air at Maximum Temperature to which it can be Preheated by Outgoing Clinker. | | |
|--|---|--|--|---|---|--|
| Exit Temperature of Kiln Gases. VII. | Number of Lbs. Steam Producing per 1 Lb. of Clinker. VIII. | Exit Temperature of Kiln Gases. IX. | Number of Lbs. Steam Producing per 1 Lb. of Clinker. X. | Number of Lbs. Steam Producing per 1 Lb. of Clinker. XI. | Temperature of Entering Combustion Air. XII. | Exit Temperature of Kiln Gases. XIII. |
| ° F. | | ° F. | | | ° F. | ° F. |
| Imposs. | Imposs. | 299.2 | ... | 0.3645 | 2039 | 917.7 |
| 266.6 | " | 382.6 | 0.0111 | 0.4919 | 1944 | 955.7 |
| 418.0 | 0.0355 | 536.3 | 0.117 | 0.4963 | 1853 | 1063.4 |
| 571.9 | 0.1467 | 692.5 | 0.2348 | 0.5742 | 1764 | 1075.1 |
| 731.9 | 0.2881 | 855.0 | 0.3884 | 0.6203 | 1674 | 1284.9 |
| 897.6 | 0.4189 | 1023.4 | 0.5441 | 0.8428 | 1584 | 1402.2 |
| 1067.5 | 0.5836 | 1195.7 | 0.6959 | 0.9856 | 1494 | 1521.8 |
| 1244.5 | 0.7755 | 1375.6 | 0.8959 | 1.155 | 1403 | 1646.1 |
| 1425.5 | 0.9900 | 1559.5 | 1.128 | 1.347 | 1315 | 1774.7 |
| 1613.5 | 1.250 | 1750.3 | 1.301 | 1.431 | 1226 | 1902.7 |
| 1808.0 | 1.553 | 1947.6 | 1.715 | 1.828 | 1137 | 2044.8 |
| 1998.1 | 1.896 | 2140.0 | 2.075 | 2.118 | 1047 | 2173.6 |
| 2007.6 | 1.922 | 2148.9 | 2.101 | 2.139 | 1041 | 2188.2 |
| 2031.0 | 1.967 | 2172.9 | 2.149 | 2.169 | 1031 | 2196.6 |
| 2047.6 | 2.013 | 2189.5 | 2.185 | 2.206 | 1024 | 2206.6 |
| 2065.1 | 2.037 | 2206.6 | 2.225 | 2.235 | 1015 | 2217.3 |
| 2089.2 | 2.086 | 2231.7 | 2.274 | 2.278 | 1006 | 2236.2 |
| 2109.6 | 2.127 | 2252.4 | 2.319 | 2.296 | 997 | 2240.2 |
| 2116.6 | 2.152 | 2258.8 | 2.343 | 2.330 | 988 | 2250.2 |
| 2124.4 | 2.180 | 2266.0 | 2.321 | 2.353 | 979 | 2251.1 |
| 2145.4 | 2.228 | 2287.2 | 2.412 | 2.388 | 970 | 2265.8 |
| 2154.1 | 2.257 | 2295.4 | 2.454 | 2.409 | 962 | 2268.5 |
| 2174.3 | 2.301 | 2315.9 | 2.499 | 2.474 | 953 | 2282.4 |
| 2195.1 | 2.351 | 2337.0 | 2.550 | 2.492 | 944 | 2297.1 |
| 2213.6 | 2.388 | 2355.8 | 2.598 | 2.561 | 935 | 2309.4 |
| 2236.9 | 2.448 | 2379.5 | 2.651 | 2.573 | 926 | 2326.4 |
| 2258.0 | 2.497 | 2401.0 | 2.707 | 2.617 | 917 | 2341.3 |
| 2279.4 | 2.549 | 2422.6 | 2.764 | 2.665 | 910 | 2357.7 |
| 2286.0 | 2.582 | 2428.7 | 2.799 | 2.686 | 899 | 2365.1 |
| 2306.8 | 2.635 | 2449.8 | 2.848 | 2.731 | 891 | 2371.6 |
| 2329.1 | 2.695 | 2472.3 | 2.906 | 2.781 | 881 | 2386.6 |
| 2349.7 | 2.749 | 2493.3 | 2.966 | 2.826 | 872 | 2401.0 |
| 2566.3 | 3.386 | 2713.4 | 3.640 | 3.367 | 781 | 2552.6 |
| 2767.0 | 4.163 | 2915.6 | 4.455 | 4.013 | 694 | 2689.1 |
| 2986.3 | 5.228 | 3137.9 | 5.573 | 4.919 | 605 | 2840.9 |
| 3162.8 | 6.672 | 3315.3 | 6.448 | 6.008 | 516 | 2950.0 |
| 3398.7 | 8.736 | 3554.3 | 9.262 | 7.819 | 428 | 3115.7 |

TABLE

Showing the Number of Lbs of Steam at 150 Lbs Pressure by Gauge per Square Retained Constant at 20 per Cent, but the Entering Air Air Supply — 10 478 lbs per 1 lb of standard coal of 12 600 B Th U s

| Clinker Production | | Entering Combustion Air Preheated to 100° F | | Entering Combustion Air Preheated to 400° F | |
|--|--|--|--|--|--|
| Tons of Standard Coal per 100 Tons of Clinker I | Lbs of Clinker per 1 Lb of Standard Coal II | Exit Temperature of Kiln Gases III | Number of Lbs Steam Producible per 1 Lb of Clinker IV | Exit Temperature of Kiln Gases V | Number of Lbs Steam Producible per 1 Lb of Clinker VI |
| 11 893 | 8 408 | °F Corresponding to entering air at 100° F | | °F | |
| 12 525 | 7 984 | | | | |
| 13 210 | 7 570 | | | | |
| 13 957 | 7 165 | | | | |
| 14 791 | 6 761 | | | | |
| 15 726 | 6 359 | | | | |
| 16 770 | 5 963 | | | 423 3 | 0 0573 |
| 17 960 | 5 568 | 426 4 | 0 0572 | 589 7 | 0 2155 |
| 19 301 | 5 181 | 594 0 | 0 2293 | 761 4 | 0 4011 |
| 20 855 | 4 795 | 770 1 | 0 4310 | 942 2 | 0 6124 |
| 22 660 | 4 413 | 954 8 | 0 6723 | 1131 9 | 0 8340 |
| 24 801 | 4 032 | 1144 1 | 0 9538 | 1325 3 | 1 1819 |
| 25 044 | 3 993 | 1159 0 | 0 9806 | 1340 1 | 1 217 |
| 25 278 | 3 956 | 1180 1 | 1 013 | 1361 6 | 1 253 |
| 25 517 | 3 919 | 1198 5 | 1 0458 | 1380 7 | 1 300 |
| 25 760 | 3 882 | 1218 2 | 1 0783 | 1400 5 | 1 324 |
| 26 021 | 3 843 | 1238 8 | 1 113 | 1421 6 | 1 376 |
| 26 274 | 3 806 | 1258 5 | 1 1483 | 1441 8 | 1 398 |
| 26 532 | 3 769 | 1272 0 | 1 1763 | 1454 6 | 1 423 |
| 26 802 | 3 731 | 1284 7 | 1 2006 | 1466 9 | 1 452 |
| 27 078 | 3 693 | 1307 3 | 1 2384 | 1489 9 | 1 496 |
| 27 352 | 3 656 | 1319 6 | 1 267 | 1501 6 | 1 522 |
| 27 632 | 3 619 | 1339 7 | 1 3003 | 1522 3 | 1 563 |
| 27 917 | 3 582 | 1360 0 | 1 342 | 1543 6 | 1 606 |
| 28 209 | 3 545 | 1377 4 | 1 357 | 1561 0 | 1 650 |
| 28 514 | 3 507 | 1401 3 | 1 422 | 1585 6 | 1 700 |
| 28 827 | 3 469 | 1422 6 | 1 466 | 1607 4 | 1 738 |
| 29 137 | 3 432 | 1433 0 | 1 506 | 1628 4 | 1 784 |
| 29 455 | 3 395 | 1456 1 | 1 537 | 1640 9 | 1 817 |
| 29 771 | 3 359 | 1476 8 | 1 574 | 1662 1 | 1 861 |
| 30 120 | 3 320 | 1498 5 | 1 626 | 1685 0 | 1 914 |
| 30 460 | 3 283 | 1520 0 | 1 673 | 1706 5 | 1 997 |
| 31 305 | 2 915 | 1737 4 | 2 194 | 1929 8 | 2 534 |
| 39 246 | 2 549 | 1954 9 | 2 673 | 2151 1 | 3 326 |
| 45 766 | 2 185 | 2187 6 | 3 517 | 2389 5 | 4 800 |
| 54 824 | 1 824 | 2390 1 | 4 828 | 2595 0 | 5 843 |
| 70 126 | 1 466 | 2650 5 | 6 581 | 2861 5 | 7 35 |

III

Inch Produccible per 1 Lb. of Clinker Produced when the Slurry Moisture is is Preheated to Various Degrees by the Outgoing Clinker

per lb. entering at 60° F. Clinkering temperature assumed to be 2,500° F.

| <i>Entering Combustion. Air Preheated to 800° C.</i> | | <i>Entering Combustion. Air Preheated to 1,000° F.</i> | | <i>Entering Air at Maximum Temperature to which it can be Preheated by Outgoing Clinker.</i> | | |
|--|---|--|--|--|---|---|
| Exit Tempera- ture of Kiln Gases. VII. | Number of Lbs. Steam Produccible per 1 Lb. of Clinker. VIII. | Exit Tempera- ture of Kiln Gases. IX. | Number of Lbs. Steam Produccible per 1 Lb. of Clinker. X. | Number of Lbs. Steam Produccible per 1 Lb. of Clinker. XI. | Tempera- ture of Entering Combustion Air. XII. | Exit Tempera- ture of Kiln Gases. XIII. |
| ° F. | | ° F. | | | ° F. | ° F. |
| ... | ... | ... | ... | ... | 2039 | 338.0 |
| ... | ... | ... | ... | ... | 1944 | 457.0 |
| ... | ... | ... | ... | ... | 1853 | 574.0 |
| ... | ... | ... | ... | 0.2690 | 1704 | 687.8 |
| ... | ... | 427.0 | 0.0519 | 0.3255 | 1674 | 796.5 |
| 477.8 | 0.2405 | 586.5 | 0.1965 | 0.4972 | 1584 | 913.8 |
| 641.3 | 0.2544 | 752.7 | 0.3565 | 0.6358 | 1494 | 1036.1 |
| 813.2 | 0.4361 | 927.9 | 0.5519 | 0.7965 | 1403 | 1164.5 |
| 991.5 | 0.6458 | 1109.4 | 0.7490 | 0.9803 | 1315 | 1298.8 |
| 1179.0 | 0.8909 | 1300.1 | 1.031 | 1.2117 | 1226 | 1455.4 |
| 1375.5 | 1.182 | 1500.2 | 1.337 | 1.447 | 1137 | 1586.9 |
| 1574.5 | 1.5205 | 1702.1 | 1.692 | 1.731 | 1047 | 1732.2 |
| 1588.1 | 1.548 | 1715.3 | 1.722 | 1.733 | 1041 | 1741.7 |
| 1611.2 | 1.5899 | 1739.1 | 1.765 | 1.795 | 1031 | 1760.4 |
| 1630.2 | 1.606 | 1758.1 | 1.805 | 1.828 | 1024 | 1773.5 |
| 1651.1 | 1.6659 | 1779.4 | 1.844 | 1.858 | 1015 | 1789.0 |
| 1673.0 | 1.710 | 1801.7 | 1.891 | 1.896 | 1006 | 1805.7 |
| 1693.9 | 1.7456 | 1822.9 | 1.932 | 1.930 | 997 | 1820.9 |
| 1705.8 | 1.778 | 1834.4 | 1.965 | 1.954 | 988 | 1826.9 |
| 1717.4 | 1.808 | 1845.6 | 1.994 | 1.972 | 979 | 1832.2 |
| 1741.1 | 1.856 | 1869.7 | 2.042 | 2.011 | 970 | 1850.3 |
| 1752.1 | 1.886 | 1880.3 | 2.071 | 2.034 | 962 | 1855.8 |
| 1773.4 | 1.928 | 1901.8 | 2.120 | 2.073 | 953 | 1871.4 |
| 1795.1 | 1.978 | 1924.1 | 2.170 | 2.116 | 944 | 1887.7 |
| 1813.5 | 2.019 | 1942.7 | 2.210 | 2.157 | 935 | 1900.6 |
| 1839.0 | 2.072 | 1968.8 | 2.271 | 2.197 | 926 | 1920.3 |
| 1861.7 | 2.123 | 1991.8 | 2.323 | 2.240 | 917 | 1937.5 |
| 1883.4 | 2.173 | 2014.0 | 2.375 | 2.283 | 910 | 1955.0 |
| 1895.0 | 2.208 | 2025.2 | 2.415 | 2.307 | 899 | 1959.0 |
| 1917.2 | 2.254 | 2047.7 | 2.468 | 2.350 | 891 | 1976.2 |
| 1940.2 | 2.315 | 2071.1 | 2.525 | 2.398 | 881 | 1992.8 |
| 1963.1 | 2.369 | 2094.4 | 2.584 | 2.447 | 877 | 2010.0 |
| 2194.5 | 3.108 | 2330.0 | 3.557 | 2.976 | 781 | 2181.8 |
| 2421.1 | 3.777 | 2559.3 | 4.058 | 3.624 | 694 | 2348.6 |
| 2667.2 | 4.832 | 2809.3 | 5.176 | 4.510 | 605 | 2530.8 |
| 2876.8 | 6.162 | 3021.1 | 6.569 | 5.646 | 516 | 2675.7 |
| 3151.7 | 8.319 | 3299.9 | 8.846 | 7.408 | 428 | 2881.4 |

TABLE

Showing the Number of Lbs of Steam at 150 Lbs Pressure by Gauge per Square
Retained Constant at 30 per Cent, but the Entering Air
Air Supply — 10 478 lbs per 1 lb of standard coal of 12,600 B Th U s

| Clinker Production | | Entering Combustion Air Preheated to 100° F | | Entering Combustion Air Preheated to 400° F | |
|--|--|--|--|--|--|
| Tons of Standard Coal per 100 Tons of Clinker I | Lbs of Clinker per 1 Lb of Standard Coal II | Exit Temperature of Kiln Gases III | Number of Lbs Steam Producible per 1 Lb of Clinker IV | Exit Temperature of Kiln Gases V | Number of Lbs Steam Producible per 1 Lb of Clinker VI |
| | | ° F | | ° F | |
| 11 893 | 8 408 | | | | |
| 12 525 | 7 984 | | | | |
| 13 210 | 7 570 | | | | |
| 13 957 | 7 165 | | | | |
| 14 791 | 6 761 | | | | |
| 15 726 | 6 359 | | | | |
| 16 770 | 5 963 | | | | |
| 17 960 | 5 568 | | | | |
| 19 301 | 5 181 | | | 366 6 | 0 01129 |
| 20 855 | 4 795 | 396 6 | 0 03692 | 544 4 | 0 2146 |
| 22 660 | 4 413 | 570 6 | 0 2592 | 726 2 | 0 4591 |
| 24 801 | 4 032 | 760 4 | 0 5312 | 920 7 | 0 7659 |
| 25 044 | 3 993 | 778 0 | 0 5596 | 938 0 | 0 7845 |
| 25 278 | 3 956 | 799 5 | 0 5739 | 950 6 | 0 8203 |
| 25 517 | 3 919 | 816 9 | 0 6209 | 978 1 | 0 8521 |
| 25 760 | 3 882 | 836 7 | 0 6533 | 998 6 | 0 8869 |
| 26 021 | 3 843 | 857 5 | 0 6877 | 1020 0 | 0 9224 |
| 26 274 | 3 806 | 877 3 | 0 7187 | 1040 4 | 0 9581 |
| 26 532 | 3 769 | 894 9 | 0 7479 | 1059 8 | 0 9933 |
| 26 802 | 3 731 | 917 4 | 0 7874 | 1079 8 | 1 036 |
| 27 078 | 3 693 | 951 6 | 0 8121 | 1094 6 | 1 060 |
| 27 352 | 3 656 | 947 0 | 0 8416 | 1109 7 | 1 089 |
| 27 632 | 3 619 | 967 0 | 0 8781 | 1130 3 | 1 128 |
| 27 917 | 3 582 | 987 8 | 0 9163 | 1152 1 | 1 170 |
| 28 209 | 3 545 | 1004 7 | 0 9487 | 1169 0 | 1 205 |
| 28 514 | 3 507 | 1030 2 | 0 9787 | 1195 3 | 1 252 |
| 28 827 | 3 469 | 1051 4 | 1 034 | 1217 1 | 1 298 |
| 29 137 | 3 432 | 1072 7 | 1 075 | 1239 2 | 1 341 |
| 29 455 | 3 395 | 1088 7 | 1 109 | 1254 7 | 1 376 |
| 29 771 | 3 359 | 1109 7 | 1 150 | 1276 3 | 1 420 |
| 30 120 | 3 320 | 1132 2 | 1 195 | 1300 0 | 1 471 |
| 30 460 | 3 283 | 1154 1 | 1 240 | 1322 0 | 1 517 |
| 34 305 | 2 915 | 1379 8 | 1 749 | 1554 5 | 2 073 |
| 39 246 | 2 549 | 1614 8 | 2 396 | 1794 9 | 2 767 |
| 45 766 | 2 185 | 1867 0 | 3 263 | 2053 9 | 3 704 |
| 54 824 | 1 824 | 2098 0 | 4 369 | 2289 5 | 4 900 |
| 70 126 | 1 466 | 2225 2 | 5 608 | 2424 6 | 6 280 |

IV

Inch Produisible per 1 Lb. of Clinker Produced when the Slurry Moisture is Preheated to Various Degrees by the Outgoing Clinker

per lb. entering at 60° F. Clinkering temperature assumed to be 2,500° F.

| Entering Combustion. Air Preheated to 800° F. | | Entering Combustion. Air Preheated to 1,000° F. | | Entering Air at Maximum Temperature to which it can be Preheated by Outgoing Clinker. | | |
|--|--|--|--|---|---|---------------------------------|
| Exit Temperature of Kiln Gases. | Number of Lbs. Steam Produisible per 1 Lb. of Clinker. | Exit Temperature of Kiln Gases. | Number of Lbs. Steam Produisible per 1 Lb. of Clinker. | Number of Lbs. Steam Produisible per 1 Lb. of Clinker. | Temperature of Entering Combustion Air. | Exit Temperature of Kiln Gases. |
| VII. | VIII. | IX. | X. | XI. | XII. | XIII. |
| ° F. | | ° F. | | | ° F. | ° F. |
| ... | ... | ... | ... | ... | 2039 | ... |
| ... | ... | ... | ... | ... | 1944 | ... |
| ... | ... | ... | ... | ... | 1853 | ... |
| ... | ... | ... | ... | ... | 1764 | ... |
| ... | ... | ... | ... | ... | 1674 | ... |
| ... | ... | ... | ... | 0.08897 | 1584 | 452.2 |
| ... | ... | ... | ... | 0.2210 | 1494 | 571.7 |
| 395.9 | 0.0370 | 494.6 | 0.1426 | 0.3644 | 1403 | 658.4 |
| 565.6 | 0.2310 | 662.2 | 0.3435 | 0.5460 | 1315 | 832.3 |
| 751.1 | 0.4681 | 856.9 | 0.6216 | 0.7743 | 1226 | 992.6 |
| 960.5 | 0.6198 | 1070.1 | 0.8867 | 1.020 | 1137 | 1146.1 |
| 1141.3 | 1.070 | 1254.2 | 1.235 | 1.236 | 1047 | 1280.9 |
| 1158.1 | 1.101 | 1270.8 | 1.266 | 1.299 | 1041 | 1294.2 |
| 1179.9 | 1.142 | 1290.5 | 1.308 | 1.337 | 1031 | 1311.8 |
| 1199.9 | 1.176 | 1313.4 | 1.343 | 1.364 | 1024 | 1327.1 |
| 1221.2 | 1.215 | 1335.2 | 1.385 | 1.399 | 1015 | 1343.9 |
| 1243.4 | 1.254 | 1357.8 | 1.427 | 1.432 | 1006 | 1361.4 |
| 1264.6 | 1.295 | 1379.4 | 1.469 | 1.467 | 997 | 1377.6 |
| 1283.8 | 1.323 | 1398.5 | 1.508 | 1.495 | 988 | 1391.6 |
| 1303.1 | 1.369 | 1417.5 | 1.545 | 1.529 | 979 | 1405.5 |
| 1317.4 | 1.401 | 1433.5 | 1.584 | 1.555 | 970 | 1416.1 |
| 1333.4 | 1.434 | 1448.0 | 1.615 | 1.573 | 962 | 1426.2 |
| 1354.9 | 1.479 | 1469.8 | 1.660 | 1.616 | 953 | 1442.6 |
| 1377.1 | 1.523 | 1492.5 | 1.708 | 1.655 | 944 | 1460.0 |
| 1395.0 | 1.578 | 1510.7 | 1.753 | 1.688 | 935 | 1472.9 |
| 1422.4 | 1.609 | 1538.5 | 1.809 | 1.735 | 926 | 1495.2 |
| 1445.1 | 1.666 | 1561.9 | 1.859 | 1.772 | 917 | 1509.5 |
| 1468.1 | 1.716 | 1585.2 | 1.909 | 1.823 | 910 | 1522.2 |
| 1483.1 | 1.752 | 1600.0 | 1.949 | 1.849 | 899 | 1540.5 |
| 1505.5 | 1.802 | 1622.8 | 2.002 | 1.892 | 891 | 1558.6 |
| 1529.6 | 1.857 | 1647.5 | 2.059 | 1.936 | 881 | 1577.0 |
| 1553.0 | 1.911 | 1671.2 | 2.117 | 1.983 | 872 | 1595.2 |
| 1794.8 | 2.525 | 1917.9 | 2.510 | 2.499 | 781 | 1783.4 |
| 2042.4 | 3.294 | 2169.2 | 3.568 | 3.106 | 694 | 1976.0 |
| 2311.0 | 4.329 | 2442.7 | 4.661 | 4.024 | 605 | 2184.8 |
| 2552.8 | 5.657 | 2687.6 | 6.046 | 5.107 | 516 | 2364.0 |
| 2698.9 | 6.933 | 2839.3 | 7.720 | 6.003 | 428 | 2443.4 |

TABLE

Showing the Number of Lbs of Steam at 150 Lbs Pressure by Gauge per Square Retained Constant at 40 per Cent, but the Entering Air Air Supply—10.478 lbs per 1 lb of standard coal of 12,600 B Th U's

| Clinker Production | | Entering Combustion Air Preheated to 100° F | | Entering Combustion Air Preheated to 400° F | |
|--|--|--|--|--|--|
| Tons of Standard Coal per 100 Tons of Clinker I | Lbs of Clinker per 1 Lb of Standard Coal II | Exit Temperature of Kiln Gases III | Number of Lbs Steam Producible per 1 Lb of Clinker IV | Exit Temperature of Kiln Gases V | Number of Lbs Steam Producible per 1 Lb of Clinker VI |
| | | ° F | | ° F | |
| 11 893 | 8 408 | | | | |
| 12 525 | 7 984 | | | | |
| 13 210 | 7 570 | | | | |
| 13 957 | 7 165 | | | | |
| 14 791 | 6 761 | | | | |
| 15 726 | 6 359 | | | | |
| 16 770 | 5 963 | | | | |
| 17 960 | 5 568 | | | | |
| 19 301 | 5 181 | | | | |
| 20 855 | 4 795 | Exit Temperature Corre sponding to IV | | Exit Temperature Corre sponding to VII | |
| 22 660 | 4 413 | | | | |
| 24 801 | 4 032 | 367 8 | 0 0030 | 506 9 | 0 2160 |
| 25 044 | 3 993 | 387 0 | 0 0324 | 526 0 | 0 2466 |
| 25 278 | 3 956 | 405 5 | 0 0615 | 545 3 | 0 2771 |
| 25 517 | 3 919 | 424 0 | 0 0898 | 564 2 | 0 3085 |
| 25 760 | 3 882 | 444 0 | 0 1216 | 584 8 | 0 3439 |
| 26 021 | 3 843 | 464 0 | 0 1538 | 605 5 | 0 3768 |
| 26 274 | 3 806 | 483 5 | 0 1863 | 625 5 | 0 4132 |
| 26 532 | 3 769 | 501 7 | 0 2158 | 643 7 | 0 4446 |
| 26 802 | 3 731 | 519 6 | 0 2463 | 661 4 | 0 4746 |
| 27 078 | 3 693 | 541 3 | 0 2816 | 683 8 | 0 5131 |
| 27 352 | 3 656 | 558 6 | 0 3116 | 700 8 | 0 5456 |
| 27 632 | 3 619 | 578 6 | 0 3443 | 721 6 | 0 5838 |
| 27 917 | 3 582 | 598 9 | 0 3822 | 742 6 | 0 6212 |
| 28 209 | 3 545 | 619 2 | 0 4181 | 763 6 | 0 6615 |
| 28 514 | 3 507 | 640 5 | 0 4573 | 785 6 | 0 7036 |
| 28 827 | 3 469 | 661 6 | 0 4959 | 807 4 | 0 7441 |
| 29 137 | 3 432 | 682 8 | 0 5350 | 829 2 | 0 7877 |
| 29 455 | 3 395 | 700 5 | 0 5686 | 847 0 | 0 8240 |
| 29 771 | 3 359 | 721 7 | 0 6091 | 868 6 | 0 8672 |
| 30 120 | 3 320 | 744 7 | 0 6552 | 891 7 | 0 9150 |
| 30 460 | 3 283 | 766 0 | 0 6948 | 914 3 | 0 9616 |
| 34 305 | 2 915 | 910 0 | 1 0135 | 1061 0 | 1 318 |
| 39 246 | 2 549 | 1240 0 | 1 8218 | 1402 0 | 2 1808 |
| 45 766 | 2 185 | 1506 0 | 2 672 | 1676 3 | 3 099 |
| 54 824 | 1 824 | 1763 0 | 3 760 | 1939 0 | 4 275 |
| 70 176 | 1 466 | 2088 0 | 5 512 | 2273 0 | 6 112 |

V

Inch Produccible per 1 Lb. of Clinker Produced when the Slurry Moisture is is Preheated to Various Degrees by the Outgoing Clinker

per lb. entering at 60° F. Clinkering temperature assumed to be 2,500° F.

| Entering Combustion. Air Preheated to 800° F. | | Entering Combustion. Air Preheated to 1,000° F. | | Entering Air at Maximum Tempera- ture to which it can be Preheated by Outgoing Clinker. | | |
|---|--|--|---|---|---|--|
| Exit Tempera- ture of Kiln Gases. VII. | Number of Lbs. Steam Producible per 1 Lb. of Clinker. VIII. | Exit Tempera- ture of Kiln Gases. IX. | Number of Lbs. Steam Producible per 1 Lb. of Clinker. X. | Number of Lbs. Steam Producible per 1 Lb. of Clinker. XI. | Tempera- ture of Entering Combustion Air. XII. | Exit Tempera- ture of Kiln Gases XIII. |
| ° F. | | ° F. | | | ° F. | ° F. |
| ... | ... | ... | ... | ... | 2039 | ... |
| ... | ... | ... | ... | ... | 1944 | ... |
| ... | ... | ... | ... | ... | 1853 | ... |
| ... | ... | ... | ... | ... | 1764 | ... |
| ... | ... | ... | ... | ... | 1674 | ... |
| ... | ... | ... | ... | ... | 1584 | ... |
| ... | ... | Exit Temperature Corre- sponding to XIII. | | ... | 1494 | Exit Temp. Corre- sponding to XIV. |
| Exit Temperature Corre- sponding to XI. | | | | ... | 1403 | |
| | | 408.9 | 0.0601 | 0.0121 | 1315 | 374.8 |
| 501.5 | 0.1988 | 595.4 | 0.3363 | 0.2234 | 1226 | 524.6 |
| 698.1 | 0.5153 | 796.0 | 0.6676 | 0.4344 | 1137 | 661.0 |
| 717.1 | 0.5456 | 814.9 | 0.7086 | 0.7048 | 1047 | 819.1 |
| 737.6 | 0.5802 | 836.6 | 0.7410 | 0.7339 | 1041 | 835.3 |
| 757.7 | 0.6163 | 856.4 | 0.7761 | 0.7655 | 1031 | 852.4 |
| 778.4 | 0.6512 | 877.6 | 0.7761 | 0.7949 | 1024 | 867.6 |
| 800.0 | 0.6888 | 899.6 | 0.8163 | 0.8266 | 1015 | 885.2 |
| 821.1 | 0.7273 | 921.1 | 0.8566 | 0.8613 | 1006 | 902.8 |
| 838.9 | 0.7638 | 921.1 | 0.8963 | 0.8945 | 997 | 919.5 |
| 856.6 | 0.8006 | 938.9 | 0.9307 | 0.9248 | 988 | 932.9 |
| 879.8 | 0.8405 | 956.5 | 0.9669 | 0.9550 | 979 | 946.0 |
| 896.8 | 0.8752 | 980.1 | 1.012 | 0.9857 | 970 | 965.0 |
| 918.5 | 0.9131 | 997.1 | 1.046 | 1.017 | 962 | 977.9 |
| 940.1 | 0.9528 | 1019.2 | 1.091 | 1.047 | 953 | 995.1 |
| 962.1 | 0.9952 | 1041.2 | 1.136 | 1.087 | 944 | 1012.9 |
| 985.0 | 1.040 | 1063.7 | 1.183 | 1.125 | 935 | 1031.0 |
| 1008.0 | 1.087 | 1087.0 | 1.231 | 1.163 | 926 | 1049.1 |
| 1030.0 | 1.038 | 1110.0 | 1.280 | 1.204 | 917 | 1068.0 |
| 1048.0 | 1.183 | 1134.0 | 1.327 | 1.246 | 910 | 1086.9 |
| 1071.0 | 1.232 | 1151.0 | 1.369 | 1.274 | 899 | 1099.0 |
| 1095.0 | 1.284 | 1174.0 | 1.419 | 1.316 | 891 | 1117.4 |
| 1118.0 | 1.337 | 1199.0 | 1.476 | 1.361 | 881 | 1136.5 |
| 1270.0 | 1.736 | 1223.0 | 1.531 | 1.406 | 872 | 1136.0 |
| 1625.0 | 2.654 | 1376.0 | 1.951 | 1.717 | 781 | 1260.0 |
| 1910.0 | 3.537 | 1739.0 | 2.946 | 2.546 | 694 | 1565.2 |
| 2182.0 | 4.902 | 2030.0 | 4.017 | 3.398 | 605 | 1795.4 |
| 2524.0 | 7.014 | 2304.0 | 5.377 | 4.475 | 516 | 2008.0 |
| | | 2660.0 | 7.618 | 6.261 | 428 | 2291.0 |

CHAPTER XXVII

THE DESIGN OF WASTE-HEAT BOILERS FOR CEMENT ROTARY KILNS

Synopsis.—A description is given of recent improvements in heat transmission through tubes, whereby the efficient recovery of waste heat has been made possible under circumstances formerly regarded as presenting insuperable difficulties.

Introduction.—The steadily increasing cost of fuel has caused the attention of industrialists to be directed to an increasing extent to the recovery of waste heat from furnaces.

The writer during recent years has been engaged in carrying out much elaborate engineering research work on boiler design for the efficient recovery of waste heat from furnaces (with special reference to wet process cement kilns with exit temperatures of 700° to 800° F.). An account of the results attained must be interesting to all manufacturers who possess furnaces from which hot gases escape thermally unutilised.

The practical upshot of this research work was to show that such hot gases could be economically utilised to a far greater extent than has hitherto been deemed possible by our manufacturers. The scientific result was to uphold the validity of the film theory of thermal resistance, first put forward by Osborne Reynolds in 1874, and later extended by the recent brilliant research work of Langmuir, Fry, Jordan, Nusselt, and other workers in America, whose researches must be regarded as of an epoch-making character from the point of view of the steam engineer.

It is now established beyond question that, by taking advantage of these recent technical advances, waste-heat boilers have been erected of an efficiency far exceeding anything generally thought possible. It is possible to use them economically with exit gases issuing at temperatures as low as 600° F., and when employed in some types of furnaces they will yield more than enough power to run the whole of the works. Dirty and dusty flue gases present no terrors for a properly designed waste-heat boiler of this type. They need practically no attention, but run month after month, with monotonous regularity.

Systems of Waste-Heat Recovery.—Before proceeding to describe the means whereby these recent technical advances have been achieved it will be advisable to give an account of the usual methods employed in waste-heat recovery.

Two main systems are in use to-day, namely: (1) the Water-Tube System, and (2) the Fire-Tube System.

Typical plant of each system will be described briefly.

1. *Water-Tube System.*—In boilers of this type tubes full of water are immersed slantwise in a chamber through which the hot flue gases pass. The water in the tubes is thereby heated to the boiling-point, and owing to the slanting position of the tubes, the hot water immediately rushes through the tubes into a reservoir above them and raises the mass of water therein to boiling. One of the best waste-heat boilers of this type is the *Edge-Moor Waste-Heat Boiler*, shown in diagrammatic view in Fig. 1. A, A, A are the water tubes set slantwise in flues through which the waste gases from the furnace pass as indicated. The water

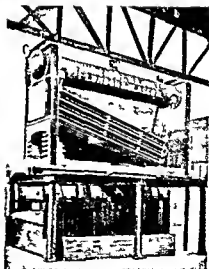
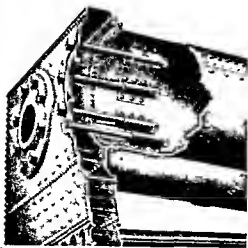
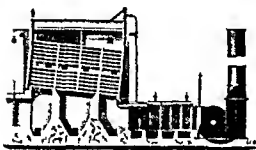
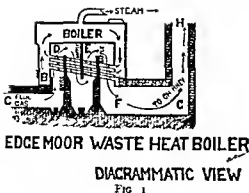


Fig 2 is a section showing the arrangement of dampers and flue

Fig 3 shows the method of staying the headers in the boiler

Fig 4 shows a boiler being erected

Fig 5 shows a series of large boilers in position in a gas works

in the tubes A, A is heated to boiling and passes out into the headers B, the hot water circulating through the boiler as shown.

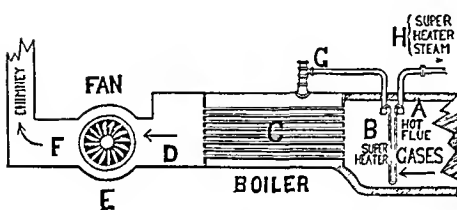
Another excellent boiler of the same type is the well-known Babcock & Wilcox boiler. Here the headers are replaced by tubes, as indicated in Fig. 6.

2. *The Fire-Tube System.*—In this system the waste flue gases are caused to traverse pipes of relatively narrow bore running through a boiler. Fig. 7 shows the plant in diagrammatic form. A is the flue; B the superheater; C the boiler; E the fan leading to the chimney, F.

Fig. 8 shows a Kirke fire-tube boiler, twin drum type, which has been used for generating steam from the waste gases of a steel furnace.

All these boilers are fitted with an automatic feed water regulator and a high and low water alarm, so that they require practically no attention.

The principle of the fire-tube waste-heat boiler is that of the well-known locomotive boiler, which has been worked successfully since the time of George Stephenson over a century ago. So that no novelty as regards principle can be claimed for this type of boiler. Nevertheless, this ancient boiler, when used for waste-heat recovery, can be so designed and so worked as far to exceed any other



MARTIN-TAYLOR WASTEHEAT BOILER PLANT.

DIAGRAMMATIC VIEW.

FIG. 7.

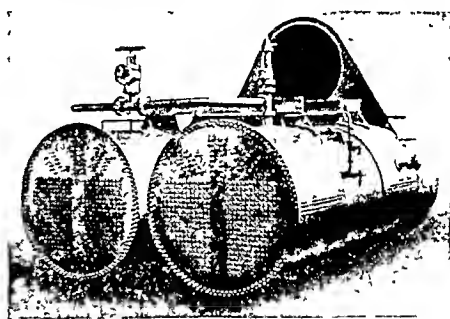


FIG. 8.

known type of boiler as regards thermal efficiency, simplicity of operation, and cheapness of installation.

This is due to the fact that advantage can be taken of the increased heat transmissibility attained by causing the hot gases to traverse the internal pipes at a very great speed. To see this needs a brief discussion of the results of modern research on heat transmission.

Comparison of the Water and Fire Tube Systems for Heat Recovery.

—As regards the *fire-tube system* the peculiar advantages it presents are as follows:—

1. The gases can be sent through the tubes at a very great speed of 80 to 100 miles per hour, thereby reducing the thickness of the gas film adhering to the steel surfaces to very small dimensions and thus effecting an extremely efficient heat transfer.
2. The plant is very simple and much cheaper to install and work than the water-tube system.
3. The plant needs practically no attention, working year in and year out with the greatest regularity and efficiency.
4. The great speed of the gases through the fire-tubes keeps dust from depositing in them. The plant is therefore particularly successful in cement works and similar works where dirty flue gases are a feature.
5. Cold air leakages can be completely avoided.

The disadvantage of the fire tube system is that very large units are difficult to construct on account of the shell thickness required

Water Tube System—The disadvantages of the *water tube system* for waste-heat recovery are obvious —

1 The relatively slow velocity of the flue gas over the boiler tubes does not destroy the thermally resisting gas film. Hence the heat transmission is bad as compared with the fire tube system

2 It follows that a very large heating surface is necessary to secure the requisite heat transmission. This makes the plant much larger and more expensive to install than the fire tube system

3 With dusty flue gases continual cleaning and great attention is necessary to secure efficiency

4 Cold air leakage is difficult to prevent. If much fan draught is put on the boiler, the brick setting leaks and allows the entry of much cold air. This reduces the quantity of steam produced, and adds uselessly to the volume of the products and consequently to the power absorbed by the fan

There can therefore be no doubt that, so far as waste heat recovery is concerned, a fire tube system is economically a better proposition than a water tube system

Effect of Gas and Water Films on Heat Transfer—In most systems

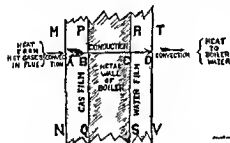


DIAGRAM SHOWING METHOD OF HEAT TRAVEL
THROUGH A WASTE HEAT BOILER WALL

FIG 9

of waste heat recovery the heat is transmitted from the hot flue gases, through a steel boiler wall, to the water, as indicated in Fig 9. The heat flows mainly by conduction, radiation playing only a small part in the process owing to the low temperature of the waste gases and the poor radiating power of hot gases

It has been proved that a stationary film of gas, MNQP, adheres firmly to the side of the steel wall which is in contact with the hot flue gas, while a stationary film of water, RSTV, coats that side of the steel plate which is in contact with the water. These films

are caused by the intense molecular attraction of the molecules forming the boundary walls of the steel plate for the surrounding gaseous or fluid molecules

The thickness of the films ranges from $1/16$ to $3/16$ in. Their effect is to coat the steel plate with two non-conducting layers which impede enormously the passage of heat through the walls from flue gas to water. In fact, the thermal resistance of the gas and water films so immensely exceeds that of the metal wall that the effect of the latter is quite negligible

This is evident from the fact that the thermal conductivities (k) of the various layers are as follows —

| | Conductivity k | | | | |
|--------------------|------------------|--------|-------|--------|---------------------------|
| <i>Air film</i> | 0.014 | B.T.U. | 's/hr | /sq ft | $^{\circ}$ F/ft thickness |
| <i>Water film</i> | 0.340 | " | " | " | " |
| <i>Steel plate</i> | 25.000 | " | " | " | " |

Since the thermal resistances are the reciprocals of the thermal conductivities, it follows that the thermal resistances of equal thicknesses of the three layers work out as 1703 : 24 : 1. Consequently, the air film opposes a resistance to the flow of heat no less than 24 times that of a film of water of equal thickness, and 1,703 times that offered by an equally thick metal wall

In other words, the only influence arresting the flow of heat which the practical engineer need consider is the air film. It is necessary to go into some detail on this matter.

Formula for Calculating Flow of Heat through a Compound Wall.—Consider a wall composed of an air film, a steel wall, and a water film, as shown in Fig. 10.

Let t_1, t_2, t_3, t_4 , be the temperatures in ° F. of the successive faces of the air film, metal wall, and water film, as indicated in Fig. 10.

Let d_1, d_2, d_3 , be the thickness of each layer in feet.

Let k_1, k_2, k_3 , be the thermal conductivity of each layer in B.Th.U.'s/hr./sq. ft./° F./ft. thickness.

Let r_1, r_2, r_3 , be the thermal resistance* of each layer, and R that of the compound wall composed of these layers.

Let A_1, A_2, A_3, A_4 , be the successive cross-sectional areas of each layer through which the same quantity of heat, Q B.Th.U.'s, flows in the same time, θ hours.

The thermal force or pressure driving this quantity of heat, Q , in succession through each layer of the compound wall, against its thermal resistance r , is the *temperature difference* which exists on the two sides of each layer. The flow of heat follows the same laws as electrical flow or the flow of a fluid through a pipe.

Hence :

$$\text{rate of heat flow across each layer} = \frac{\text{thermal driving force across layer}}{\text{thermal resistance of layer}},$$

$$\text{or} \quad \frac{Q}{\theta} = \frac{t_1 - t_2}{r_1} = \frac{t_2 - t_3}{r_2} = \frac{t_3 - t_4}{r_3} \quad (1)$$

$$\text{hence} \quad \frac{Q}{\theta} = \frac{(t_1 - t_2) + (t_2 - t_3) + (t_3 - t_4)}{r_1 + r_2 + r_3},$$

$$\text{or} \quad \frac{Q}{\theta} = \frac{t_1 - t_4}{r_1 + r_2 + r_3} \quad (2)$$

$$\text{or} \quad \frac{Q}{\theta} = \frac{t_1 - t_4}{R} \quad (3)$$

The thermal resistance, R , of the compound wall is the sum of the individual resistances r_1, r_2, r_3 , of the successive layers composing the wall. Further, each individual resistance, r , is proportional to the thickness, d , of the layer, and inversely proportional to the cross-sectional area, A , of the path through which the heat flows and inversely as the specific thermal conductivity of each layer which makes up the wall.

$$\text{Hence} \quad r_1 = \frac{d_1}{A_1 k_1} \quad r_2 = \frac{d_2}{A_2 k_2} \quad r_3 = \frac{d_3}{A_3 k_3} \quad (4)$$

$$\text{and} \quad R = r_1 + r_2 + r_3 = \frac{d_1}{A_1 k_1} + \frac{d_2}{A_2 k_2} + \frac{d_3}{A_3 k_3} \quad (5)$$

* Unit thermal resistance is that resistance which allows a unit quantity of heat to flow in unit time through a unit area when a unit difference in temperature is maintained across a unit thickness. The thermal resistance is the reciprocal of the thermal conductivity.

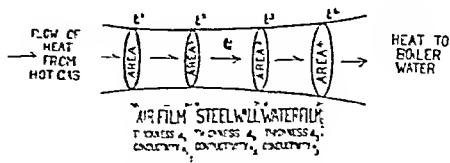


DIAGRAM SHOWING HOW HEAT FLOWS
THROUGH A COMPOUND WALL

FIG. 10.

Substituting this value of R in (3), we get

$$Q = \frac{t_1 - t_4}{\frac{d_1}{A_1 k_1} + \frac{d_2}{A_2 k_2} + \frac{d_3}{A_3 k_3}} \quad (6)$$

From this formula most practical problems relating to the flow of heat through a compound wall can be solved

Practical Application of Formula to Waste-heat Boiler—Now let us apply formula (6) to the practical case of a *waste heat boiler*

Make $A_1 = A_2 = A_3 = A_4 = 1$ sq ft

Substitute the conductivities —

For air film $k_1 = 0.014$ B Th U /hr /sq ft / 1° F /1 ft thick

For steel boiler plate $k_2 = 25,000$ B Th U 's „ „ „

For water film $k_3 = 0.34$ B Th U „ „ „

Let d_1, d_2, d_3 be the thickness (in feet) of the air film, steel plate, and water film successively, and t_1 the temperature of the *flue gas* and t_4 the temperature of the *boiler water* in $^\circ$ F

Then we get by substituting in (6)

B Th U 's transmitted per hour per square foot of heating surface

$$= \frac{t_1 - t_4}{\frac{d_1}{0.014} + \frac{d_2}{25,000} + \frac{d_3}{0.34}}$$

$$\text{or} \quad \text{B Th U s/hr /sq ft} = \frac{t_1 - t_4}{71.4d_1 + 2.94d_2 + 0.017d_3} \quad (7)$$

From this formula we can calculate the heating surface required to produce a given supply of steam provided we know the temperature of the flue gas and the thickness of the various films which coat the boiler plate

In a practical case let $t_1 - t_4 = 1,000^\circ$ F

For the air and water film let the thickness

$$d_1 = d_3 = 0.1 \text{ in} = 0.00833 \text{ ft}$$

For the boiler plate let the thickness

$$d_2 = 0.5 \text{ in} = 0.0417 \text{ ft}$$

Then (7) becomes

$$\text{B Th U 's/hr /sq ft} = \frac{1000}{0.5950 + 0.0244 + 0.0017} = 1,610 \text{ B Th U 's}$$

The main influence stopping the flow of heat is the resistance of the gas film. A long way behind this in potency is the water film, while of quite negligible importance is the thickness and nature of the metal enclosing the water

Increase of Efficiency Caused by Reducing the Thickness of the Gas Film on the Boiler Plates—It is not possible to reduce the thickness of the gas film to zero. It is, however, possible to reduce its thickness very considerably. The following table (p. 277) shows what an enormous effect the reduction of the thickness of the gas film has on increasing the flow of heat from the flue gases into the water per square foot of heating surface.

Reducing the thickness of the gas film by one half increases the flow of heat by nearly 100 per cent. By reducing the thickness of the gas film to one tenth its original amount, we increase the flow of heat by over 600 per cent, and by reducing the thickness to one hundredth of its original amount, we increase

TABLE I

| Thickness of Gas Film . . . | 0.1 in. | 0.05 in. | 0.01 in. | 0.001 in. | 0.0001 in. | Zero. |
|---|---------|----------|----------|-----------|------------|--------|
| <i>B.Th.U.'s transmitted per 1 sq. ft. of heating surface per hour per 1,000° F. temperature difference .</i> | 1610 | 3090 | 11,682 | 31,250 | 37,453 | 38,314 |

the flow of heat from 1,610 to 31,250 B.Th.U.'s per sq. ft., or by nearly 2,000 per cent.

The effect of reducing the thickness of the water film is negligible in comparison with the effect of reducing the gas film :—

TABLE II

| Thickness of Water Film. | B.Th.U.'s Transmitted per Square Foot of Heating Surface per Hour per 1,000° F. Temp. Difference. |
|--------------------------|---|
| In. | |
| 0.1 | 1610 |
| 0.05 | 1642 |
| 0.01 | 1669 |
| 0.001 | 1675 |
| 0.0001 | 1676 |

Here you will see that reducing the thickness of the water film to one-half its original thickness only increases the flow of heat by 2 per cent., and reducing the thickness to one-thousandth part of the original thickness merely increases the flow of heat by 4 per cent. Reducing the thickness of the metal shell of the boiler to zero merely increases the flow of heat by 0.25 per cent.—a perfectly negligible amount.

Hence we now see what we must do in order to increase the efficiency of a waste-heat boiler. *We must reduce the thickness of the gas film covering the steel boiler plate to the smallest practical dimensions possible.*

Method of Reducing the Thickness of the Gas Film by Increasing the Speed of Gaseous Flow over the Boiler Plate.—We have just seen that the predominant influence checking the flow of heat into the boiler water from the flue gas is the presence of a gas film. We have also shown that even a slight reduction in the thickness of this film enormously increases the efficiency of the boiler by increasing the transmission of heat.

The problem now is: How are we to reduce the thickness of this gas film? The answer is: By increasing the speed of gaseous flow over the metal surfaces. By this means we tear away many of the molecules of the gas adhering to the metal, and thus reduce the thickness of the stationary gaseous layer.

Let us consider this effect in greater detail. Consider a gas flowing along a steel boiler tube AAAA (Fig. 11). The gas sorts itself out into different layers moving at different speeds. In the centre occurs a core of gas (GG) moving at the maximum speed. Surrounding it occurs a more slowly moving layer FF.

Along the sides is a layer of stationary molecules, forming the heat resisting gaseous film

The gaseous molecules attract each other. Consequently, one layer of molecules moving rapidly over another tend to pull or drag the slower moving molecules along with them. The more rapid the motion of the gaseous layer, the more pronounced is this dragging action. The more rapidly the gas flows down the pipe, the greater will be the relative motions between the stationary layer *EE* and the next layer *FF*, and consequently the more pronounced will be the tendency of the layer *FF* to pull away and set in motion the stationary layer *EE*. It is therefore obvious that the movement of furnace gases over the heating surfaces tends to destroy or make thinner the gas film clinging to the plates, and the greater the velocity of the gas, the thinner the gas film.

We should therefore expect that increasing the speed of gas through a fine tube will greatly increase the rate of heat transmission through such a tube. Experiment has confirmed this.

In 1844 Peclet pointed out the existence of a water film ("Traité de la Chaleur," chap viii, p 13). In 1874 Osborne Reynolds (*Proc Man Lit & Phil Soc*, 1874, p 9) published an epoch making paper, in which he put forward

the idea that the transmission of heat depended upon the speed of the gas. Stanton (*Phil Trans Roy Soc* vol 190, p 67) and Nicholson confirmed this work.

In 1912 Langmuir, in America, began to publish his remarkable researches on gas films and the effect of velocity upon the thickness of the film and upon heat transfer (*Phys Review* 34 p 421, *Trans Amer Electrochem Soc*, 23 299, 1913). He demonstrated by means of a delicate thermocouple that on a hot silver plate between 150° and 500° C an air film of 4 mm existed.

DIAGRAM SHOWING PASSAGE OF FLUE GAS THROUGH A BOILER TUBE

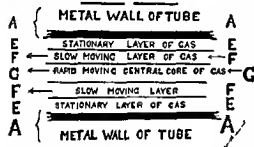


FIG 11

The work was continued in America and an extremely important series of researches has been published there within the last few years which has settled the question, so far as ordinary engineering is concerned, of the heat transfer of flue gases to boiler water*.

Lolog Formulæ for Heat Transmission—As this work is so fundamentally important, I will give a brief account of it.

The preceding authors had clearly established the fact that when hot flue gases pass down a boiler tube the heat transmission in *B Th U* s into the water per unit of time and per unit of heating surface is not the same for each unit of tube length.

The curve (Fig 12) shows this. The ordinates represent the amount of heat remaining in each lb of waste gases for each unit length along the tube. The temperature difference between the water in the boiler and the waste gases is larger at the commencement of the tube than at the end of the tube. This causes a much greater amount of heat to be transmitted into the water in the earlier parts of the tube than towards its end. Thus by the time one half of the tube

* The main references are Wm Nusselt *Verein Deutsche Eng* 89 1910 Fessenden and Haney *University of Missouri Bull* 27 No 26 (1916) H P Jordan *Proc Inst Mech Eng* 1909 Fry *Trans A S M E* 1917

has been traversed the gas has yielded up to the water about 75 per cent. of its available heat.

Professor Hedrick showed that the loss of heat from a gas travelling down a flue is given accurately by the expression :—

$$\text{Log } H_x = c \cdot e^{-x} \quad (8)$$

$$\text{or} \quad -\frac{dH_x}{dx} = cmH_x e^{-mv} \quad . \quad . \quad . \quad . \quad . \quad (9)$$

where x is the length of the tube traversed by the gas, c and m are constants, and H_x is the quantity of heat (reckoned from absolute zero) remaining in the gas after it has traversed the length x , while e is the Napierian base of logarithms.

The physical significance of this is that the heat in the gas is imparted to the water at a rate which diminishes exponentially as the distance x increases down the tube, but at any one particular distance x the rate is directly proportional to the quantity of heat H_* left in the gas, as reckoned from absolute zero.

This formula may be applied practically to boiler design as follows :—

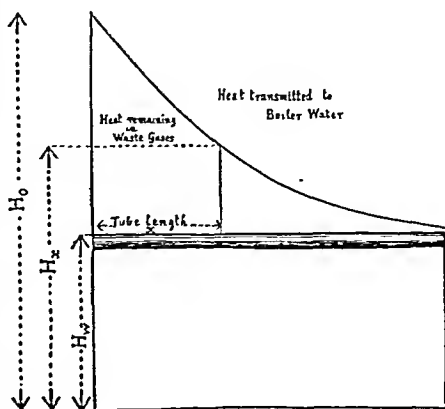


FIG. 12.

Let H_0 = absolute quantity of heat (in B.Th.U.'s) contained in 1 lb. of gas (reckoned from absolute zero) as it enters the boiler tube ($x=0$).

H_x = absolute quantity of heat (in B.Th.U.'s) remaining in 1 lb. of gas (reckoned from absolute zero) after it has traversed a length x ft. of boiler flue.

H = absolute quantity of heat (in B.Th.U.'s) contained in 1 lb. of gas (reckoned from absolute zero) when its temperature reaches the temperature of the water in the boiler.

Then in (8) let $x=0$ and e^{-mx} becomes $=1$ and $\log H_0=C$.

So that C is the quantity of heat (reckoned from absolute zero) contained in 1 lb. of gas as it enters the tube ($x=0$).

Then (8) becomes

$$\text{Log } H_x = (\text{log } H_0) e^{-mx} \quad . \quad . \quad . \quad . \quad (10)$$

Now the practical conditions of the problem require that when the tube is infinitely long ($x = \infty$) the temperature of the gas ultimately reaches that of the surrounding water, and therefore $H_x \rightarrow H_\infty$ as $x \rightarrow \infty$.

Hence, divide (10) by $\log H_n$ and we get

$$\text{Log } \frac{H_r}{H_{\infty}} = \log \frac{H_0}{H_{\infty}} e^{-mv} \quad . \quad . \quad . \quad . \quad (\text{II})$$

Equation (11) satisfies all the practical boundary conditions because $x = \infty$ gives $e^{-mx} = 0$,

giving $\text{Log } \frac{H_{\infty}}{H_{\tau}} = 0$ or $H_{\tau} = H_{\infty}$,

and

$$x=0 \text{ gives } e^{-mx}=1,$$

giving the identity

$$\text{Log } \frac{H_0}{H_\infty} = \log \frac{H_0}{H_\infty}$$

Now, taking logs of (11), we obtain the formula

$$\text{Log } \left\{ \log \frac{H_x}{H_\infty} \right\} = \log \left\{ \log \frac{H_0}{H_\infty} \right\} - mx \quad (12)$$

or, as is more usually written

$$\text{Log log } \frac{H_x}{H_\infty} = \text{log log } \frac{H_0}{H_\infty} - mx \quad (12a)$$

We must now evaluate m . It is independent of the length of tube x , and must be a coefficient depending upon the size of the pipe and the weight of gas flowing.

More heat can be abstracted from the same weight of gas if the *cross sectional* area of the passage is reduced. For this increases the number of contacts the molecules of the gas make with the surface of the tube through which it is flowing. Now the *hydraulic mean depth* (k) is a factor which is intimately related to the frequency with which a gaseous molecule will strike the surface of the wall, as is obvious from the definition

$$\text{Hydraulic mean depth of full pipe} = \frac{\text{cross sectional area}}{\text{perimeter}}$$

Hence m should contain the hydraulic mean depth (k) as a factor.

Also we have seen that the greater the *speed of gaseous* flow through a pipe, the greater the transmissibility of heat. Hence m must also contain an allowance for the *speed* of flow of the gas through the pipe.

Now it has been found empirically from the results of the experiments referred to above that

$$m = 0.193(wk)^{-0.41} \quad (13)$$

where w = weight of gas flowing in lbs/hr

k = hydraulic mean depth of flue

For a full circular pipe of diameter d in, completely filled with flowing gas we have

$$k = \frac{\pi d^2 / 4}{\pi d} = \frac{d}{4} \quad (14)$$

where d is the diameter of the tube in inches. These formulæ allow exact design of boilers to be made when the tubes are of 2 in diameter or less. In fact, so precisely do the practical results follow these laws that we can predict exactly the amount of steam producible from any given size of fire tube boiler, provided that we know the chemical composition and temperature of the exit gas. The efficiency of such boilers, designed from these formulæ, enormously exceeds that of the older types. Obviously, however, for each particular furnace the boiler must be carefully designed, the tubes put in of the right length and diameter, the fan power supplied correctly calculated from the viscosity of the flue gas, so that they can be driven through the tubes at the correct speed. The frictional resistances of the hot gas in the pipes and flues must be calculated so as to avoid the deposition of dust and eddy currents. In other words, scientific design is essential.

Calculation of the Heat Content of the Gases.—In order to obtain the heat content of the gases it is necessary to calculate the molecular specific heat of the compound gas by means of the formulæ:—

(1) For diatomic gases, air, N_2 , O_2 , H_2 , CO ,

$$C_p = 6.93 + 0.1254 \times 10^{-6} T^2 \text{ B.Th.U.'s per 1 lb. molecule per } 1^\circ \text{ F. increase.}$$

(2) For carbon dioxide, CO_2 ,

$$\text{For } T < 2900; C_p = 7.15 + 3.9 \times 10^{-3} T - 0.6 + 10^{-6} T^2.$$

$$\text{For } T < 2900; C_p = 12.196 + 0.42 \times 10^{-3} T.$$

(3) For water vapour, H_2O ,

$$C_p = 7.8 + 0.3 \times 10^{-3} T + 0.315 \times 10^{-6} T^2.$$

For ordinary boiler flue gas the mean specific heat at constant pressure (C_p) per lb.-molecule, measured from the absolute zero of temperature to T° absolute on the Fahrenheit scale, is given by

$$C_p = 6.6 + 0.0004 T \text{ B.Th.U.'s.}$$

The mean molecular weight may be taken as 28.7.

For other types of flue gases the mean specific heat must be calculated from that of the individual gases and their percentage amounts.

Practical Example of Application to Boiler Design of Lolog Formula.

—To give a practical example of the use of these formulæ in designing a boiler, suppose that we have a flue gas issuing from a furnace at a temperature of $1,500^\circ \text{ F.}$ Let the weight of flue gas to be passed through a single boiler tube be 100 lbs./hr. Let the diameter of the boiler tube be 2 in. and its length 20 ft. It is desired to calculate the heat transmission to the water through this tube in B.Th.U.'s/hr. when the temperature of the boiler water is to be 360° F. , corresponding to a steam pressure of 140 lbs./sq. in. by gauge. Knowing this, we can calculate the size of the boiler and the quantity of steam producible thereby corresponding to the given fuel consumption of the furnace.

Step 1.—Calculate $\log H_0/H_w$.

Temperature of flue gas is $1500 + 460 = 1,960^\circ \text{ F. absolute (F.)}$.

Molecular specific heat of flue gas is $(6.6 + 0.0004 \times 1960) = 6.784$.

Molecular weight is 28.7.

$$\text{Whence } H_0 = \frac{1960 \times 7.784}{28.7} = 504.3 \text{ B.Th.U.'s per 1 lb. gas.}$$

For the water temperature $T = 360^\circ + 460^\circ = 820^\circ \text{ F. absolute.}$

$$\text{Hence } H_w = 820 \times (6.6 + 0.0004 \times 820) / 28.7 \\ = 198 \text{ B.Th.U.'s per 1 lb. gas.}$$

$$\log \frac{H_0}{H_w} = \log \frac{504.3}{198} = 0.40603$$

$$\text{Lolog } \frac{H_0}{H_w} = \log 0.40603 = \bar{1}.60856.$$

Step 2 — Calculate the value of mx from (13)

$$W = 100 \text{ lbs/hr}$$

$$k = \frac{1}{4} \times 2 = \frac{1}{2}$$

whence

$$m = 0.193 \times (5)^{-0.41}$$

$$= 0.038812$$

$$x = 20 \text{ ft, whence } mx = 0.77624$$

Step 3 — Calculate H_x from (12).

$$\text{Lolog } H_x/H_w = \text{lolog } H_0/H_w - mx$$

$$= \bar{1} 60856 - 0.77624$$

$$= \bar{2} 83232$$

$$= \log (0.06797)$$

whence

$$\text{Log } H_x/H_w = 0.06797$$

or

$$H_x/H_w = 1.1695$$

whence

$$H_x = 1.1695 \times H_w = 1.1695 \times 198$$

$$= 231.5 \text{ B.Th U's per 1 lb gas}$$

Step 4 — Calculate the heat transmitted per tube

The heat transmitted per 1 lb of gas flowing is

$$H_0 - H_x = 504.3 - 231.5 = 272.8 \text{ B.Th U's/hr}$$

Hence

$$100 \text{ lbs of gas transmit } 24,780 \text{ B.Th U's/hr,}$$

which is the heat yielded per tube to the water

Step 5 — Calculate the yield of steam per tube

From our steam tables we know that with feed water at 60°F to produce 1 lb of saturated steam at 140 lbs/sq in pressure requires 1,172 B.Th U's. Hence each boiler tube will yield $27,280/1172 = 23.2$ lbs steam per hour.

Hence, knowing the total weight of flue gas available from our furnace, we have the data required to calculate the yield of steam producible, and the number of tubes which must be fitted in the boiler shell in order to give the desired steam supply.

Technical Research Work—In order to test these conclusions, elaborate research work was carried out by the writer at the invitation of the late Mr Bamber for the Associated Portland Cement Manufacturers which completely established the correctness on the large scale of the above formulæ. Nearly £5,000 was expended on this work, a model waste heat boiler of special design composed of tubes 13 ft 10 in \times 1½ in was erected at Swanscombe Cement Works in 1923, and was worked by the hot flue gases with such success that a large scale boiler was later installed, and proved most successful, giving almost theoretical results.

The larger boiler was attached to a small kiln, which produced 3½ to 4 tons of clinker per hour with a slurry containing as much as 42 per cent of moisture. With less slurry moisture the results would have been considerably more favourable, but even under these conditions, with an exit temperature of 758°F near the superheater, the boiler produced 985 lbs of steam per ton of clinker at 130 lbs per sq in, and a temperature of 485°F .

Fig 13 shows the installation, which is reproduced by kind permission of the Editor of *Cement Manufacture*, being taken from a paper on "Waste Heat Boilers," by A. C. Davis, in April 1930.

The results were so encouraging that a similar boiler was installed at the Beddington Cement Works near Croydon, while the writer has recently erected three specially designed boilers of a fire-tube type—one at the Thames Portland Cement Works at Cliffe, near Gravesend, and another at Messrs Masons' Cement Works at Claydon, near Ipswich, and another at Asheham, near Lewes.

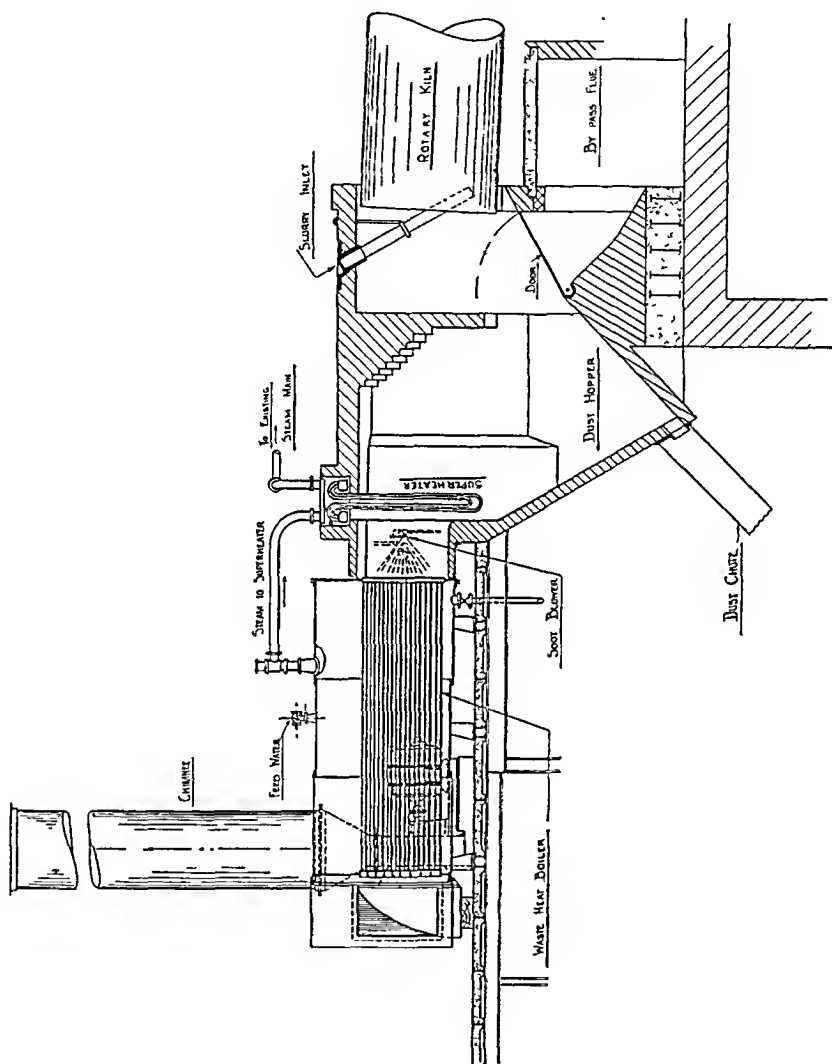


FIG. 13.—Waste-heat Fire-tube Boiler.

The results show that from cement kilns working the wet process with exit gases entering at as low a temperature as 700° F., a supply of steam can be obtained sufficient to run nearly half the works. In general, with the wet process of cement manufacture, 100 tons of cement give 50 tons of high-pressure steam.

This was formerly regarded as impossible. In fact, it was denied that a waste-

heat boiler could be used with a wet process cement plant—an absolute prejudice devoid of any basis of fact. By careful design of the flues and gaseous velocities, dust troubles are entirely eliminated, and the boilers work smoothly, month after month, with practically no attention and with automatic efficiency.

TABLE III

Typical Technical Data for the Design of a Fire-tube Waste-heat Boiler for Given Amounts of Steam for Use with Wet-process Cement Kilns

Data { *Fuel Consumption*—29 tons standard coal (of 12 600 B Th U's/lb) per 100 tons clinker
Slurry Moisture—40 per cent
Exit Temperature—800° F at entry to boiler 414° F at exit
Composition of Gas by Weight—CO₂=26.1 per cent, N₂=47.7 per cent, excess air =3.3 per cent, H₂O=22.9 per cent
1 Lb. Clinker gives 5.36 lbs. of exit gas occupying 181 cub. ft. at 800° F
Specific Heat of Exit Gas—0.297
Density of Exit Gas at 800° F, compared to air at the same temperature is 0.94,
 1 cub. ft. of gas at 800° F weighs 0.0296 lb., and at 400° F it weighs 0.0434 lb.

| Clinker Output | Volume of Exit Gas | | Weight of Exit Gas | Number of Tubes in Boilers 17 Ft. x 1½ In. | B Th U's Available in Exit Gas above 414° F | Steam at 180 Lbs Gauge and 150° F Super heat | Gross I H P at 10 Lbs of Steam per I H P | I H P of Motor Driving Fan | Available I H P for Works |
|----------------|--------------------|------------------|--------------------|--|---|--|--|----------------------------|---------------------------|
| | At 400° F | At 800° F | | | | | | | |
| Tons per Hr | Cub. Ft. per Sec | Cub. Ft. per Sec | Lbs. per Sec | | Per Sec | Lbs. per Hr | | | |
| 1 | 76.9 | 112.7 | 3.335 | 104 | 382.2 | 1.091 | 109 | 7 | 102 |
| 2 | 153.9 | 225.5 | 6.670 | 208 | 764.4 | 2.182 | 218 | 14 | 204 |
| 3 | 230.8 | 338.2 | 10.005 | 312 | 1147.0 | 3.273 | 327 | 21 | 306 |
| 4 | 307.8 | 450.9 | 13.342 | 415 | 1529.0 | 4.363 | 436 | 29 | 407 |
| 5 | 384.7 | 563.6 | 16.676 | 519 | 1911.0 | 5.454 | 545 | 36 | 509 |
| 6 | 461.6 | 676.2 | 20.014 | 623 | 2293.0 | 6.545 | 654 | 43 | 611 |
| 7 | 538.6 | 789.0 | 23.350 | 726 | 2676.0 | 7.636 | 764 | 50 | 714 |
| 8 | 615.5 | 901.7 | 26.680 | 830 | 3058.0 | 8.726 | 873 | 57 | 816 |
| 9 | 692.8 | 1015.0 | 30.020 | 934 | 3440.0 | 9.817 | 982 | 64 | 918 |
| 10 | 769.2 | 1127.0 | 33.350 | 1040 | 3822.0 | 10.910 | 1091 | 71 | 1020 |
| 11 | 846.4 | 1240.0 | 36.685 | 1141 | 4204.0 | 12.000 | 1200 | 79 | 1121 |
| 12 | 923.5 | 1353.0 | 40.020 | 1245 | 4587.0 | 13.090 | 1309 | 86 | 1223 |
| 13 | 1000.0 | 1465.0 | 43.36 | 1349 | 4969.0 | 14.180 | 1418 | 93 | 1325 |
| 14 | 1077.0 | 1578.0 | 46.69 | 1452 | 5350.0 | 15.270 | 1527 | 100 | 1427 |
| 15 | 1154.0 | 1691.0 | 50.03 | 1556 | 5733.0 | 16.360 | 1636 | 107 | 1529 |
| 16 | 1231.0 | 1803.4 | 53.36 | 1660 | 6115.0 | 17.450 | 1745 | 114 | 1631 |
| 17 | 1308.0 | 1916.0 | 56.70 | 1764 | 6497.0 | 18.545 | 1854 | 121 | 1733 |
| 18 | 1385.0 | 2029.0 | 60.04 | 1867 | 6880.0 | 19.635 | 1964 | 129 | 1835 |
| 19 | 1462.0 | 2142.0 | 63.37 | 1971 | 7262.0 | 20.725 | 2072 | 136 | 1936 |
| 20 | 1539.0 | 2255.0 | 66.70 | 2075 | 7644.0 | 21.815 | 2181 | 143 | 2038 |

Amount of Steam Obtainable from Waste-heat Boilers—The amount of steam obtainable depends largely upon the fuel consumption per 100 tons clinker produced, upon the amount of water in the slurry, and upon the amount of excess air. By varying these factors the amount of steam can be increased or diminished to almost any desired amount.

Each particular case must be worked out on its merits.

The following table shows the amount of steam produced on wet and dry process kilns working with varying amounts of water in the slurry with a *minimum* fuel consumption in the kiln for producing good clinker —

TABLE IV

| Fuel Consumption. Tons of Standard Coal of 12,600 B.Th.U.'s per 100 Tons of Clinker Made. | Slurry Moisture. Per Cent. | Per 1 Ton Clinker Made. Lbs. of Steam at 200 Lbs. by Gauge and 150° F. Superheat. | H.P. Developed at 10 Lbs. of Steam per I.H.P. Hour. | H.P. Required for Fan. | Nett H.P. Available per 1 Ton of Clinker per Hour. | Nett I.H.P. to Run Whole Works per 1 Ton per Hour Clinker Made. |
|--|----------------------------------|---|--|------------------------------|--|---|
| 28.5 | 40 | 1091 | 109 | 7 | 102 | 300 |
| 26.0 | 30 | 1208 | 121 | 7 | 114 | 300 |
| 25.0 | 20 | 1694 | 169 | 7 | 162 | 300 |
| 23.0 | 10 | 1777 | 177 | 7 | 170 | 300 |
| 21.0 | 0 | 1926 | 193 | 7 | 186 | 300 |

By burning more coal than is requisite to produce cement clinker, the amount of available heat can be immensely increased, and enough steam can be produced to more than run the whole works—as is indeed the practice in the United States.

The usual rule—which provides an ample margin—for the total H.P. required for a wet-process cement works is to multiply the weekly output by 1.8. Thus a works producing 1,000 tons of cement a week would require 1,800 I.H.P., or, per 1 ton clinker made per hour, about $168 \times 1.8 = 300$ H.P.

So that, using modern turbines consuming 10 lbs. of steam per I.H.P., we require about 3,000 lbs. of steam per 1 ton clinker made per hour, or 1.34 lbs. of steam per 1 lb. clinker made.

The following table shows the approximate fuel consumptions of kilns requisite to generate all the power necessary to run the whole works, using slurry moistures varying from 40 to 0 per cent., and keeping the excess air down to about 5 per cent. :—

TABLE V

Approximate Minimum Fuel Consumption for Cement Rotary Kilns Generating Sufficient Steam in Waste-heat Boilers to Run the Whole Works

| Fuel Consumption. Tons of Standard Coal (12,600 B.Th.U.'s per Lb.) per 100 Tons of Clinker Made. | Slurry Moisture. | Lbs. of Steam at 200 Lbs. Gauge and 150° F. Superheat per 1 Lb. of Clinker Made. | Lbs. of Steam per 1 Ton of Clinker Made. |
|--|------------------|---|--|
| | Per Cent. | | |
| 39.5 | 40 | 1.34 | 3000 |
| 35.3 | 30 | 1.34 | 3000 |
| 31.5 | 20 | 1.34 | 3000 |
| 28.6 | 10 | 1.34 | 3000 |
| 26.6 | 0 | 1.34 | 3000 |

The burning of additional coal over that required solely for cement production is obviously misleading as regards the coal consumption for clinker burning, but the practice cannot be termed wasteful, because, if the coal supply to the kiln were diminished, the steam not generated at the end of the kiln would have to be made up by supplying coal to separately fired boilers. Now, in the rotary

kiln the coal is burnt under the best conditions, the free oxygen being only about 1 per cent, so that only about 5 per cent excess air need be present, and practically no noticeable amount of carbon monoxide is produced, whereas furnaces of boilers must work with a much larger amount of excess air otherwise great waste (through carbon monoxide formation) occurs

Consequently, a waste heat boiler working in combination with a cement kiln forms ideal conditions for economical combustion, and the boiler, being so far removed from the fierce heat of the furnace, is not subjected to the wear and tear and deterioration prevalent on ordinary boilers working on separately fired furnaces

Composition of the Flue Gases from Cement Kilns—Engineers and works chemists will find useful the table on p. 27 17 showing the composition, thermal properties, and steam raising power of the exit gases issuing from kilns burning different amounts of fuel and using slurries containing different amounts of minimum moisture, all the flue gas containing amounts of excess air needed for complete combustion

Design of Flues for Cement Kiln Waste-heat Boilers—Where *waste heat* boilers are employed, it is obvious that the supply of power depends upon the regular operation of the kilns. It is necessary, therefore, that the load should be regular, night and day, seven days the week. Also, any stoppages of the kilns (which should be seldom) must not prevent output being maintained in auxiliary departments. Consequently, it is necessary to have one independently fired boiler under steam to carry on essential operations during kiln stoppages and balance the load. For the same reason, on many works the waste heat from several kilns may be poured into one flue, whence it is distributed to the boilers. The disadvantage of this latter arrangement is the enormous cost of the flues and dampers. To prevent cold air leakage through the bricks, it is practically essential to have an external steel casing, and to prevent excessive loss of heat by radiation, a layer of insulating bricks or material must be placed between the steel shell and the firebrick lining. This arrangement allows of flexibility, but some loss of heat is inevitable, and a loss of 20° to 100° F. between exit from the kiln and inlet to boilers is not unusual.

Probably the best arrangement is a separate boiler attached to each separate kiln, being cheaper and involving less loss of heat in the connecting flues. In many cases the installation of a steam accumulator would pay. A by pass flue must be provided, so that the kiln can be worked independently of the boiler in case of a breakdown of the latter.

Dust and Water Troubles with Waste-heat Boilers—In a cement works dust troubles are prevalent, but by skilful rational design of the flues and exits they can be almost entirely eliminated.

The dust adhering to the heating surfaces must be removed preferably every eight hours by means of portable steam lances of suitable design suitably installed.

In fire tube boilers the dust laden gases must never be allowed to pass directly into a cold boiler, otherwise the condensed moisture will cause caking of the flues and much trouble will result. The boiler water must invariably be preheated either by steam from other boilers or by an independent furnace before connecting up to the flue gases.

If attention is paid to these points, no trouble arises with dust, and the boilers work year in, year out, with a minimum of trouble.

Obviously, the water for waste heat boilers must be carefully softened and controlled, as scale deposits will soon ruin such boilers.

TABLE VI

Composition and Properties of Exit Gases from Dry and Wet Process Kilns Using Different Percentages of Moisture in Slurry and Different Fuel Consumptions

| Fuel Consumption. Tons of Standard Coal (of 12,600 B.Th.U.'s per Lb.) per 100 Tons of Clinker. | Slurry Moisture. | Percentage Composition of the Exit Gas by Weight 100 Lbs. | | | | Weight of Exit Gas per 1 Lb. of Clinker. | Volume of Exit Gas at Exit Tem- perature per 1 Lb. of Clinker Made. | Tempera- ture of Exit Gas. | Density of Exit Gas Air=1 at Exit Tem- perature Degree. | Per 1 Lb. of Clinker Made. B.Th.U.'s Available above 414° F. | Mean Specific Heat of Flue Gas between 414° F. and Exit Tem- perature Degree | Lbs. of Steam per 1 Lb. of Clinker Made at 200 Lbs. Gauge, 150° F. Superheat. | Lbs. of Steam per 1 Ton of Clinker Made at 200 Lbs. Gauge, 150° F. Superheat. |
|---|---------------------|--|----------------|------------------|---------|--|---|----------------------------------|--|---|--|---|---|
| | | CO ₂ | N ₂ | H ₂ O | X's Air | | | | | | | | |
| | Per Cent. | Lbs. | Lbs. | Lbs. | Lbs. | Lbs. | Cub. Ft. | ° F. | | | | | |
| 28.5 | 40 | 26.1 | 47.7 | 22.9 | 3.3 | 5.36 | 181.0 | 800 | 0.94 | 614.5 | 0.297 | 0.4753 | 1061 |
| 39.2 | 40 | 26.55 | 51.45 | 18.40 | 3.60 | 5.85 | 284.0 | 1400 | 0.96 | 1717.0 | 0.298 | 1.33 | 2975 |
| 26.0 | 30 | 30.09 | 49.75 | 16.67 | 3.49 | 4.012 | 149.6 | 1020 | 0.984 | 697.0 | 0.287 | 0.5390 | 1208 |
| 35.3 | 30 | 27.83 | 51.95 | 16.57 | 3.65 | 5.21 | 270.5 | 1600 | 0.985 | 1778.0 | 0.297 | 1.37 | 3070 |
| 25.0 | 20 | 30.88 | 50.11 | 15.48 | 3.53 | 3.798 | 171.1 | 1325 | 0.999 | 978.0 | 0.283 | 0.7564 | 1694 |
| 31.5 | 20 | 29.86 | 52.92 | 13.49 | 3.73 | 4.565 | 256.7 | 1786 | 1.005 | 1770.0 | 0.292 | 1.36 | 3047 |
| 23.0 | 10 | 34.00 | 52.83 | 9.47 | 3.70 | 3.287 | 159.5 | 1534 | 1.033 | 1026.0 | 0.279 | 0.7933 | 1778 |
| 28.5 | 10 | 32.20 | 55.36 | 8.56 | 3.88 | 3.95 | 230.7 | 1958 | 1.043 | 1726 | 0.283 | 1.34 | 2990 |
| 21.0 | 0 | 36.78 | 54.92 | 4.46 | 3.84 | 2.9131 | 167.8 | 1826 | 1.09 | 1111.0 | 0.270 | 0.8596 | 1926 |
| 26.5 | 0 | 34.11 | 56.88 | 5.01 | 3.99 | 3.577 | 224.0 | 2210 | 1.07 | 1718.0 | 0.2674 | 1.33 | 2980 |

Some Typical Waste-heat Boiler Plants—Although the installation of waste heat boilers has been possible on English works since, at least, 1906, the first plant was only put up at Swanscombe Cement Works in 1923 by the late Mr H D K Bamber, and the writer carried out the preliminary research work. This long delay was partly due to lack of enterprise rather than technical difficulties, and partly to the neglect of technical research by the British cement trade.

In the United States, however, the first plant was erected in 1902 at the Cayuga Lake Cement Co. By 1921 most American cement works were so fitted. The same applied to European, Australian, and New Zealand cement works.

These waste heat boilers provide all the steam required with efficient turbo-generators and electric drives throughout, although this necessitates (as seen above) burning somewhat more coal in the kilns than is necessary for clinker production.

Most of the plants in the United States were erected by the Edge Moor Co and the Babcock & Wilcox Co, and were of the water tube type.

Between 1926 and 1929 the Dewey Portland Cement Co, Davenport, Iowa, erected three kilns, 11×175 ft, each of capacity 250 tons per day, and each fitted to a 900 H P Edge Moor four pass boiler. This plant provides all the electrical power necessary for the works and quarries, including shovels. Here the wet process is employed.

At Oglesby, in the United States, are two kilns, 200×11 ft (producing 5,600 tons of cement per week), equipped with three 1,500-H P Edge Moor boilers, one being used as a stand by. In addition this works contains eight 1,000 H P boilers attached to kilns producing 9,100 tons of cement per week.

These boilers produce all the power necessary to run the works and quarries. The wet process is employed.

Messrs Babcock & Wilcox Ltd have specialised in waste heat boilers of the water tube type. Drawings Nos 27091/1 (Fig 14) and 27091/2 (Fig 15), 27150/16B, and 28650A are drawings of actual plant put in by this firm, the first two drawings showing a boiler of 11,976 sq ft heating surface with horizontal tube economiser, which was supplied by them to Messrs Wilson's Portland Cement Co, New Zealand, in 1925.

The sizes of the rotary kilns supplying the hot gases were

No 2 Kiln—8 ft 8 in diameter \times 160 ft

No 3 Kiln—9 ft 6 in diameter \times 8 ft \times 160 ft

The power generated from the two kilns is 1,600 to 1,700 kw. The turbo is supplied with steam at 200 lbs pressure, 100° F of superheat, and works under a vacuum of $28\frac{1}{2}$ in. The water evaporated per hour is stated to be from 22,000 to 24,000 lbs.

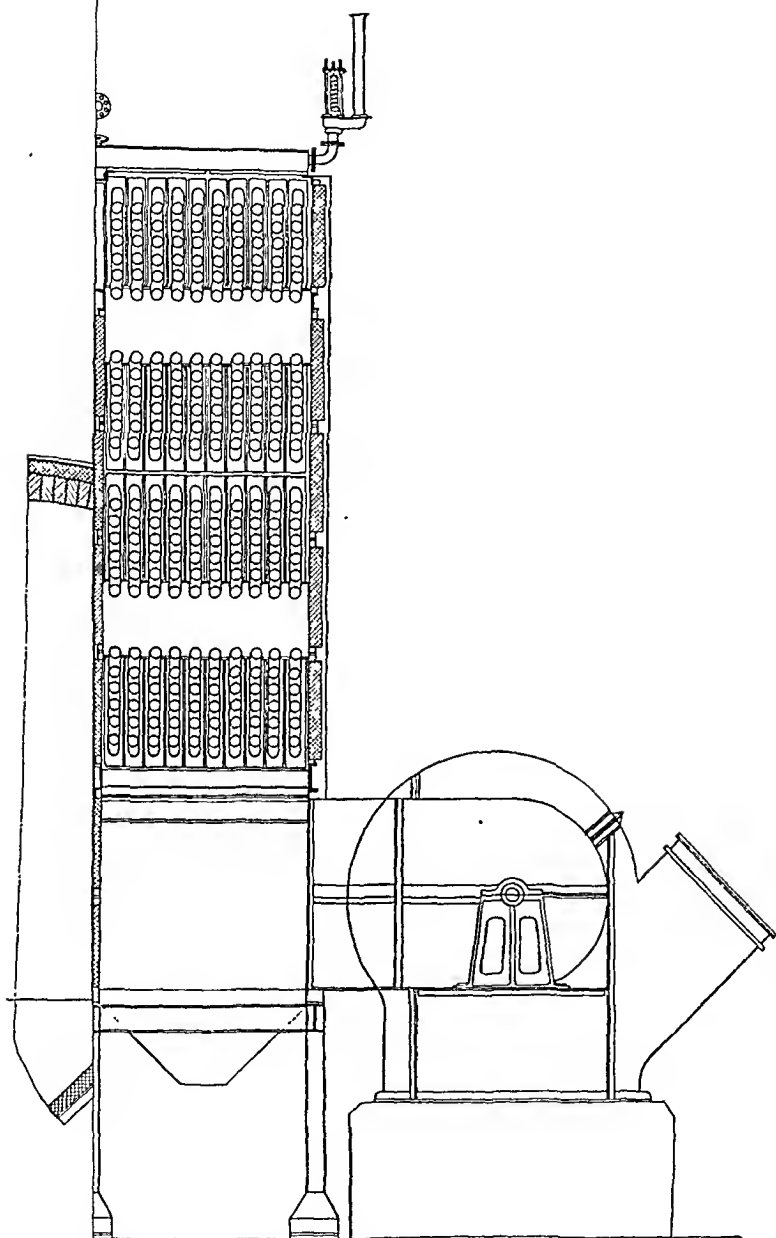
The temperatures of the gases were given as

- (a) Entering the boiler at 932° F (but it is more likely that at the boiler inlet the temperature would be $1,100^\circ$ F and the 932° F measured was close to the water tubes).
- (b) Entering the economiser at 455° F
- (c) Leaving the economiser at 350° F

The fan horse power was 112, giving a suction draught at the fan of 6 in and at the entrance to the boiler of 1.3 in.

A further boiler was installed in 1929, and the following information has been given to us.

On a trial for four weeks, with both boilers operating 1,264 hours out of a possible 1,334 hours, with three out of four rotary kilns connected with the flues, the average power developed was 2,500 kw, the steam consumption being 13.5



Zealand. The Boiler was of 11,970 Sq. Ft. Heating Surface,
12 Ft. Long.

[To face page 27.18.]

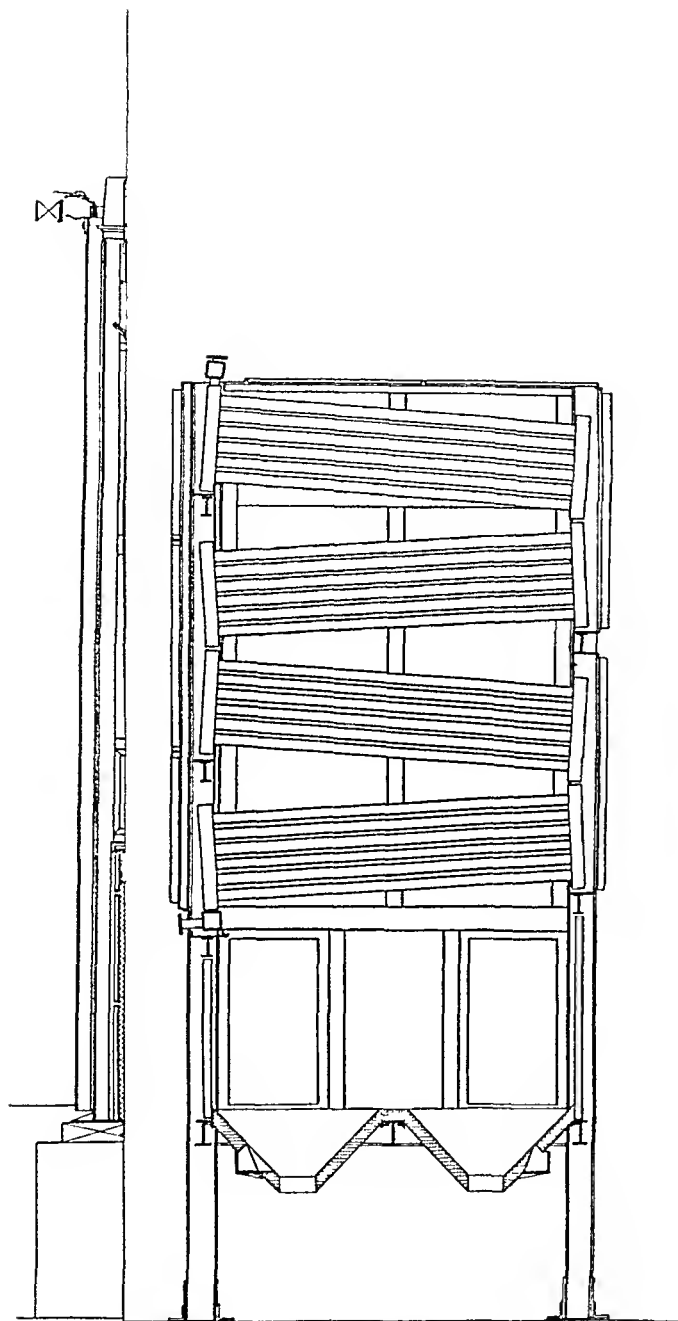


FIG. Zealand. The Boiler was of 11,976 Sq. Ft. Heating
2 Ft. long.

[To face page 27.18.]

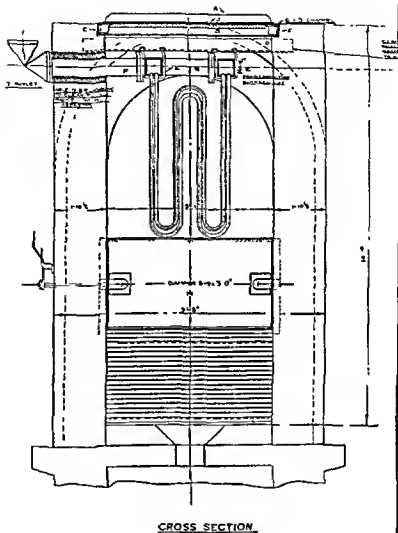


FIG. 17 (Drawing No 25705/18) —Arrangement of Si

lbs. per kw., which equals an evaporation of 36,750 lbs. per hour from three kilns.

| | |
|--|---------|
| Average temperature feed water entering economiser | 180° F. |
| Average temperature steam entering boiler | 350° F. |
| Average temperature of exit gases | 350° F. |
| Average temperature superheated steam | 500° F. |

The rotary kiln hours were :

| | |
|-------------|--------------|
| Wet process | 1,097 hours. |
| Dry process | 766 " |

Total . 1,863 hours.

The quantity of fuel burnt was 4,214 tons, and the calorific value 11,491 B.Th.U.'s per lb.

With the four kilns running, two on the dry and two on the wet process, the two boilers would give an evaporation of 47,000 lbs. of steam per hour, and with four kilns all on the wet process the evaporation would be at least 45,000 lbs. of water per hour. Each kiln output averaged 150 tons per day, and produced 833 kw. per hour.

Drawing No. 28650A (Fig. 16) shows a plant supplied by the same firm to Japan, with a horizontal superheater at the inlet. The expected performance from this boiler of 11,000 sq. ft. heating surface would be 28,000 lbs. evaporation per hour at 225 lbs. pressure, total temperature 590° F., with the gases off a kiln output of 18,000 barrels per day of twenty-four hours, dry process, gas temperature 1,500° F., feed-water temperature 100° F.

Drawing No. 25705/18 (Fig. 17) shows their latest arrangement of superheater to be placed in the inlet flue and arranged vertically with damper so that a proportion of the gases can be by-passed from the superheater. In this case the weight of gases was given to us as 46,700 kilos, at a temperature of 520° C., and the evaporation estimated to be obtained was 5,700 kilos.

A waste-heat boiler supplied to Japan was designed so that it could be used either for utilising the waste heat from the kiln on dry process, or as a pulverised fuel-fired boiler. The working pressure was 255 lbs. and the steam output superheated to a total temperature of 350° C. As pulverised fuel had to be fired, it was necessary to place the superheater interdeck.

The performance figures estimated are as follows :—

| | Weight of Gases. Kilos/hr. | Temperature. | Evaporation. Kilos/hr. |
|----------------------|----------------------------------|--------------|---------------------------|
| A ₁ . . . | 50,600 kilograms | 480° C. | 5,300 kilograms |
| A ₂ . . . | 51,400 " | 440° C. | 4,600 " |
| B ₁ . . . | 54,800 " | 485° C. | 5,700 " |
| B ₂ . . . | 55,700 " | 445° C. | 5,000 " |
| C ₁ . . . | 52,700 " | 540° C. | 6,600 " |
| C ₂ . . . | 53,500 " | 500° C. | 5,950 " |
| D ₁ . . . | 57,100 " | 545° C. | 7,000 " |
| D ₂ . . . | 58,000 " | 505° C. | 6,350 " |

Alternatives to Waste-heat Boilers.—In most cases the instalment of waste-heat boilers would pay. In cases, however, where the kilns are very efficient and work with a low exit temperature of, say, under 500° F., and where

cheap electric power can be purchased, it may be financially more remunerative to buy power than generate steam for oneself

Indeed, with the grid system of electric supply, super power stations and long distance, high tension electric transmission power can be delivered to many works so cheaply that it scarcely pays manufacturers to entail the heavy capital expenditure for power plants and the worry of running them. Also, in several cases the employment of 1,000 H P Diesel engines, using crude oil, may be more economical than electrical power, although heavy depreciation must be allowed for

Each case must be decided on its merits. It is certain, however, that all the technical difficulties attendant on the use of waste heat boilers have now been overcome. All that is needed is efficient technical design and previous experience in this type of work on the part of the designing engineers

CHAPTER XXVIII

OVERCOMING HEAT LOSSES IN CEMENT MANUFACTURE

[Tons as used in this article are English tons of 2,240 lbs. To convert the figures to American tons of 2,000 lbs. multiply by 1.12.]

Recent Research Work on Cement Kiln Design, with Special Reference to Dust, Spray, or Flotation Kilns, and Kilns with Enlarged Calcination Zones

FOR many years it has been well known that the rotary kiln, as at present designed, is extremely inefficient from a thermal point of view.

Thus an ideal kiln could yield 100 tons of cement clinker by the combustion of 6.35 tons of standard coal. In actual practice 100 tons of clinker require anything from 22 tons of coal (dry process) to 30 tons of coal (wet process); and these figures are usually immensely exceeded in most works.

Consequently many efforts have been made to construct kilns of greater thermal efficiency. A brief account may therefore be interesting of recent efforts in this direction.

Before explaining the nature of these efforts a few words must be said regarding recent discoveries concerning the *causes* of the relative inefficiency of cement kilns, because all these efforts for improvement are largely based on the discovering of what is wrong in existing kilns.

It should be mentioned that the main portion of this chapter is based upon research work carried out by the writer when director of research of the British Portland Cement Research Association, between 1922 and 1925.

Since the closing down of the Research Association in 1925 (following on the tragic death of the late H. D. K. Bamber, one of the foremost advocates for scientific research in the British cement trade) the research work was continued privately by the writer.

Requirements of a Cement Kiln

A cement kiln to be commercially successful should (1) be thermally efficient; (2) possess a large output. Thermal efficiency without output does not lead to commercial success.

For example, the shaft kiln is considerably more thermally efficient than a rotary kiln, and yet rotary kilns have replaced shaft kilns almost everywhere, except in a few countries where labour is very cheap and fuel very dear.

A good shaft kiln can, under suitable circumstances, produce 100 tons of clinker from 14 to 18 tons of standard coal (of 12,600 B.Th.U.'s per lb.) against 22 to 30 tons of coal from a rotary kiln, both working the dry process.

The disadvantage of the best shaft kilns, however, is their low output—say, 150 tons a week against 1,000 to 2,000 tons a week of a modern cement kiln, with even less labour than is requisite for a shaft kiln.

Now why is a shaft kiln more thermally efficient than a rotary kiln?

In order to explain this we must go into the thermal history of Portland cement clinker.



FIG. 1.—General View of Asheham Works where Large scale Experiment in Testing Kinds was carried out by a Group of English Financiers
Boiler House is on Top of the Hill and Raw Material Drying Plant at the Base

Thermal History of Portland Cement Formation

In order to make Portland cement we must heat a mixture of clay and calcium carbonate (chalk, limestone, etc.).

But the heat must be supplied in a definite way. We must not only supply a certain *quantity of heat*, but we must also supply it at the right thermal pressure (or temperature, as physicists say).

In order to see this, let us now follow the thermal history of a batch of raw material through the kiln.

As water plays no part in the formation of cement, and is a mere mechanical admixture, it simplifies the treatment of the subject to consider that the mixture is dry at a temperature of 60° F. (15.6° C.).

Assume that this batch of raw material consists of.

- 1.1905 lbs. of calcium carbonate (CaCO_3).
- 0.1846 lb. of kaolin ($\text{Al}_2\text{Si}_2\text{O}_7 \cdot 2\text{H}_2\text{O}$).
- 0.1540 lb. of hydrated silica (93 per cent. SiO_2 , 7 per cent. H_2O).
- 0.0312 lb. of ferric oxide (Fe_2O_3).

Total, 1.5603 lbs.

This mixture, when ignited to about $2,498^{\circ}$ F. (1370° C.), will form 1.00 lb. of cement clinker of the ordinary manufacture in Great Britain.

As the temperature (or thermal pressure) of this 1 lb. mass of raw material increases—*i.e.*, as it becomes hotter and hotter—different quantities of heat are absorbed; or, so to speak, are pumped into the material at different temperature levels (or better, at different thermal pressures).

The accompanying Table I. shows this, which I have calculated from data compiled by the British Portland Cement Research Association.

TABLE I

Quantity of Heat Absorbed by the Solid Raw Materials in Making 1 Lb. of Portland Cement Clinker between 32° F. and the Clinkering Temperature $2,498^{\circ}$ F.

| t° F. | Quantity of heat in B.Th.U.'s. | |
|----------------|---|---|
| 32 | 0.00 | |
| 212 | { 70.29 at beginning of 212° F. 77.09 at end of 212° F. | { Due to splitting off of water } 70.29 B.Th.U.'s in drying zone. |
| 312 | 117.60 | |
| 412 | 158.04 | |
| 512 | 198.59 | |
| 612 | 239.02 | |
| 712 | 279.57 | |
| 812 | 320.01 | |
| 912 | 360.54 | |
| 1012 | 400.89 | |
| 1112 | { 441.42 at beginning of 1112° F. 455.32 at end of 1112° F. | { Due to splitting off of water from kaolin } 534.20 B.Th.U.'s in preheating zone. |
| 1212 | 495.68 | |
| 1312 | 536.08 | |
| 1412 | 576.47 | |
| 1481 | { 604.49 at beginning of 1481° F. 1416.41 at end of 1481° F. | { 811.92 B.Th.U.'s absorbed owing (to evolution of CO_2 by CaCO_3) } Decarbon- ating zone. |

TABLE I—(continued)

| $^{\circ}\text{F}$ | Quantity of heat in B Th U s | |
|--------------------|---|---|
| 158r | 1444 61 | Sintering zone } 286 70 B Th U s absorbed in heating the raw materials up to 2498°F and 179 93 B Th U s are evolved when chemical union takes place making net absorption 10 677 B Th U s |
| 168r | 1472 81 | |
| 178r | 1501 01 | |
| 188r | 1529 21 | |
| 198r | 1557 41 | |
| 208r | 1585 61 | |
| 218r | 1613 71 | |
| 228r | 1641 9r | |
| 238r | 1670 21 | |
| 248r | 1698 31 | |
| 249S | { 1703 11 at beginning of 2498°F } Due to exothermic { 1523 18 at end of 2498°F } reaction | |

Necessary to Force Different Quantities of Heat at Different Thermal Pressures

By studying this table you will see that between 32° and 212°F the raw material must have forced into it a quantity of heat measured by 70 29 B Th U s so that from a practical point of view 70 29 B Th U s must be pumped into the raw material at a thermal pressure not below 21°F

Next between 212° and 1481°F you will see that the material must have forced into it an additional 534 B Th U s so that the kiln designer must arrange matters that this number of B Th U s must be available from the surrounding heating medium at a thermal pressure between 212° and 1481°F

Next at 1481°F —the temperature at which the calcium carbonate begins to decompose and evolve carbon dioxide—the material must have pumped into it no less than 812 B Th U s delivered at a thermal pressure above 1481°F

If these B Th U s are available from the heating medium at a lower thermal pressure than 1481°F they are of no practical value so far as decomposing the calcium carbonate and expelling the CO_2 therefrom is concerned. At a lower thermal pressure than 1481°F the B Th U s simply cannot be pumped into or absorbed by the material and they pass away through the kiln unabsorbed and without doing any useful chemical work and escape up the chimney to the air unutilised merely increasing the exit temperature of the kiln gases in so doing.

Finally about 107 B Th U s must be forced into the material between 1581° and 2498°F in order to cause the lime and silica to chemically unite to form the mixture of calcium silicates known as Portland cement clinker. And if these 107 B Th U s are not available from the surrounding heating medium at a thermal pressure of about 2500°F then no cement clinker is formed no matter how many millions of B Th U s are forced through the kiln or how many tons of coal are burned.

This fact was strikingly illustrated some years ago by costly large scale experiments carried out by the Associated Portland Cement Manufacturers at their Swanscombe Works by Eldred. He obtained hot gas derived from a producer, and lowered the temperature of the gas below 2500°F by mixing exit gas with it and forced the mixed gases through a rotary kiln fed with slurry. The material came out merely underburnt and the exit temperature at the end of the kiln shot up, while the fuel consumption increased enormously.

These experiments cost nearly £5 000 and demonstrated the impossibility of making cement by these methods.

Heat Available above 1,481° F. Valuable in Cement Clinker Formation

It will be seen, then, that the temperature of 1,481° F. (805° C.) is a critical temperature at which a great deal of heat is absorbed in the cement kiln at a very high thermal pressure or temperature, because it is the temperature at which the calcium carbonate decomposes.

Hence the kiln may be conveniently divided into two parts—a portion *A*, where the temperature of the raw material is above 805° C. (1,481° F.), and a portion *B*, where the temperature of the raw material is *below* 1,481° F.

So that in making 1 lb. of cement clinker the raw-material mixture must be subjected to a steadily increasing temperature, starting at 60° F. and finishing at about 2,498° F., and the dry raw mixture must absorb 652.5 B.Th.U.'s below 1,481° F. and 918.6 B.Th.U.'s *above* 1,481° F. *before* 1 lb. of cement clinker can be formed.*

Since 1,481° F. (808° C.) is the temperature whereat the calcium carbonate begins to decompose in the furnace, obviously no clinker can be formed until the chalk *does* decompose. So that any heat escaping unutilised past the line dividing the *A* portion of the kiln from the *B* portion is completely lost so far as cement formation is concerned—the heat thus escaping being merely employed in pre-heating and drying the raw material preparatory for its conversion into clinker, but the bulk of it escapes unutilised up the chimney, as just explained above. The quantity of clinker formed, therefore, is measured by the amount of heat absorbed by the raw material between 1,481° F. (805° C.) and the clinkering temperature, 2,498° F. (1,370° C.).

But in a cement kiln practically all heating is effected by means of the hot gas evolved from burning coal or oil. Consequently this quantity of heat, absorbed by the raw material between 2,498° and 1,481° F., must in the ideal case be equal to the amount of heat given out by the hot gas between the time of entering the kiln portion *A* and leaving it for the portion *B*, since it is the hot gas which is the heating agent.

Now 1 lb. of standard coal (12,600 B.Th.U.'s per lb.) yields 11.278 lbs. of hot combustion gas. If this combustion gas from 1 lb. of coal enters into the section *A* of the kiln at its maximum temperature, *T*° F., and leaves the section *A* at the temperature of 1,481° F. (805° C.), the amount of heat given up by the 11.278 lbs. of combustion gas passing down the section *A* is:

$$Q = 11.278 \times S \times (T - 1481^\circ) \text{ in B.Th.U.'s}$$

where *S* is the mean specific heat of the combustion gas.

But the amount of heat required to be absorbed by the raw material in making 1 lb. of clinker between 1,481° F. (805° C.) and the clinkering temperature of 2,498° F. (1,370° C.) is 918.6 B.Th.U.'s. Therefore, we get the weight in lbs. of clinker *W* formed by the 11.278 lbs. of hot gas derived from 1 lb. of standard coal in the section *A* as:

$$W = \frac{11.278 \times S \times (T - 1481^\circ)}{918.6} \text{ lbs.}$$

* It should be carefully noted that it is a fallacy to state that the total minimum quantity of heat required to make 1 lb. of cement clinker is 652.5+918.6=1571.1 B.Th.U.'s, because this statement takes no account of the quantity of heat recoverable from the hot clinker and evolved gases. Heat is, so to speak, pumped *into* the raw material from the external surroundings, but much of this heat flows, so to speak, *through* the material and escapes back at a lower temperature level (or thermal pressure) to the external surroundings again. In much the same way that a stream of high-pressure water flowing through a turbine escapes at a low pressure after having done its work—the work being done by loss of pressure—so also, in cement formation, the main amount of work is done by the B.Th.U.'s losing their temperature or thermal pressure, the quantity of heat being not very greatly altered thereby.

Connection between Flame Temperature and Clinker Output

In the accompanying Table II I have calculated the number of B Th U 's liberated by the 11,278 lbs of combustion gas evolved from 1 lb of burning coal, when it cools from a given temperature T° to $1,481^{\circ}$ F, and the number of lbs of clinker which can be produced therefrom theoretically, also the number of tons of standard coal (of 12,600 B Th U 's per lb) consumed per 100 tons of clinker produced.

The highest possible flame temperature theoretically attainable by this gas is $5,370^{\circ}$ F ($2,966^{\circ}$ C), which occurs when the combustion air is preheated by the escaping clinker to $2,500^{\circ}$ F ($1,371^{\circ}$ C).

A study of this table shows forcibly the correctness of the contention put forward above—namely, that it is not only the *quantity of heat* which counts, but even more so the thermal pressure (or *temperature*, as physicists call it), at which the heat is delivered to the substance.

For example, 100 B Th U 's of heat available at a thermal pressure of 212° F are much less valuable than 100 B Th U 's available at, say, $2,600^{\circ}$ F.

TABLE II

Connection between Flame Temperature and Clinker Output

| Temperature of Gas Derived from Com- bustion of 1 Lb of Standard Coal of 12,600 B Th U 's per Lb | | Number of B Th U 's Liberated by 11,278 Lbs of Combustion Gas Falling from T° to $1,481^{\circ}$ F (805° C) | Number of Lbs of Clinker producible per 1 Lb of Standard Coal Burnt | Tons of Standard Coal Consumed per 100 Tons of Clinker Produced |
|---|--------------|--|---|--|
| $^{\circ}$ F | $^{\circ}$ C | B Th U 's | Lbs Clinker | Tons Coal |
| 5370 | 2966 | 14,452 | 15 73 | 6 36 |
| 4100 | 2266 | 9 301 | 10 125 | 9 88 |
| 3000 | 1649 | 5,115 | 5 568 | 17 96 |
| 2900 | 1594 | 4,759 | 5 181 | 19 30 |
| 2800 | 1538 | 4,405 | 4 795 | 20 85 |
| 2700 | 1482 | 4 053 | 4 413 | 22 66 |
| 2600 | 1427 | 3,704 | 4 032 | 24 80 |
| 2500 | 1371 | 3,358 | 3 656 | 27 35 |
| 2400 | 1315 | 3,016 | 3 283 | 30 46 |
| 2300 | 1260 | 2 677 | 2 915 | 34 30 |
| 2200 | 1204 | 2,341 | 2 549 | 39 25 |
| 2100 | 1149 | 2,007 | 2 185 | 45 77 |
| 2000 | 1093 | 1 676 | 1 824 | 54 82 |
| 1900 | 1038 | 1,347 | 1 466 | 70 13 |
| 1800 | 982 | 1,021 | 1 111 | 90 00 |
| 1700 | 926 | 698 | 0 760 | 131 58 |
| 1600 | 871 | 378 | 0 411 | 243 31 |
| 1500 | 815 | 60 | 0 065 | 1538 46 |
| 1481 | 805 | 0 | 0 00 | Infinite |

In the one case 100 units of heat will produce no cement clinker at all, and in the other case the 100 units of heat can be converted into an equivalent amount of clinker.

The case is analogous to the thermal phenomena involved in the formation of steam in a boiler. For example, millions of B Th U 's passed through the heating

flues of a boiler at a temperature below the boiling point of water (100° C. or 212° F.) will not generate steam. The elevation of the temperature of the heating gases above 212° F. is what decides the value of the heating medium so far as steam raising is concerned.

In the same way, in cement production, the value of the gaseous heating medium for producing cement clinker is measured by the number of B.Th.U.'s available *above* 805° C. ($1,481^{\circ}$ F.).

For the sake of convenience, the B.Th.U.'s available *above* 805° C. ($1,481^{\circ}$ F.) are called "high-grade" heat, and B.Th.U.'s available *below* 805° C. are called "low-grade" heat.

It will also be seen that the ordinary thermal balance of the rotary kiln is practically valueless from a technical point of view, because the quantities of heat are all expressed as so many B.Th.U.'s without distinguishing whether these B.Th.U.'s are available at a high or low thermal pressure or temperature.

For example, the table shows us that when the flame temperature is only $1,481^{\circ}$ F., it is necessary to burn an infinite amount of coal in order to produce an ounce of clinker, whereas, when the flame temperature is at its maximum of $5,370^{\circ}$ F. ($2,966^{\circ}$ C.), we can produce 100 tons of cement clinker by the expenditure of only 6.36 tons of standard coal. With a flame temperature of $2,600^{\circ}$ F. we could produce 100 tons of clinker by the combustion of 24.8 tons of standard coal—which is practically obtained in many cases.

It is no exaggeration to say that ignorance of this factor has been largely responsible for the lack of progress in cement-kiln design and construction during the last thirty years. Many costly experiments have been carried out which (it could have been predicted beforehand) were doomed to failure because the designers were unaware of this fundamental fact of thermodynamics—for example, Eldred's experiments mentioned above.

Radiation Losses of the Thermal Units Available at High Thermal Pressures (or Temperatures) are Main Causes of Cement Kiln Inefficiencies

We are now in a position to discuss the question: Why is a shaft kiln so much more efficient than a rotary kiln?

The main reason is undoubtedly due to the greater external and internal radiation losses in one case than in the other. It is the radiation losses which are the main cause of inefficiency.

We have just shown that in making 1 lb. of cement clinker it was necessary for the raw material to absorb from the surrounding hot gaseous heating agent 652.5 B.Th.U.'s *below* 805° C. ($1,481^{\circ}$ F.) and 918.6 B.Th.U.'s *above* 805° C., so that in order to make 1 lb. of cement clinker, 652.5 low-grade B.Th.U.'s and 918.6 high-grade B.Th.U.'s are essential.

There is no difficulty in supplying the low-grade B.Th.U.'s required for cement making. The supply of the high-grade B.Th.U.'s is, however, a much more difficult matter, as heat is extremely rapidly lost by radiation and convection at temperatures exceeding 805° C. ($1,481^{\circ}$ F.).

The rate at which heat losses by radiation occur increase according to Steffan's law in proportion to the fourth power of the absolute temperature. So that when the temperature becomes very high, enormous quantities of heat are lost in this way by radiation from hotter to colder surfaces. If we *double* the absolute temperature, we increase the rate at which high-grade heat is lost 16 times. If we *treble* the absolute temperature, the heat is lost 81 times as fast! So that when temperatures become very high the number of high-grade B.Th.U.'s lost by radiation at these high temperatures becomes perfectly enormous.

It will be seen, therefore, that the really valuable B.Th.U.'s are lost by the

gaseous heating medium above 805°C ($1,481^{\circ}\text{F}$), which are precisely those liable to be lost by radiation to colder bodies in the neighbourhood

To every 918.6 high grade B Th U 's lost in this way there corresponds the loss of 1 lb of potential clinker

Extent of Clinker Losses by External Radiation

When we examine the question of the thermal losses by external radiation from the rotary kiln in the light of this conception, we find that the effect of the external radiation losses are much more serious than has heretofore been believed

For example, if we measure the percentage of external radiation losses from the rotary kiln, measured on the total number of B Th U 's liberated in the kiln by the coal, we find that about an 8 per cent loss occurs

When however, we come to work these losses out in terms of high grade and low grade B Th U 's we find that a much higher proportion of high grade B.Th U 's are lost than of low grade, because the bulk of the radiation occurs from the hot or clinkering, end of the kiln

Accurate experimental measurements showed that, owing to this cause, the loss of *clinker formation* due to external radiation amounted to nearly 15 per cent, which is a serious loss

In other words, in a rotary kiln making 100 tons of clinker, if we stopped completely all external radiation losses we would make 115 tons of clinker for the same fuel consumption as before

Enormous Clinker Losses Due to Internal Radiation

It is, however, when we come to consider the loss of heat by *internal radiation* that this question of losses becomes paramount

The rotary kiln, it must be remembered, is a tube with one end red hot and the other end comparatively cold. High grade heat is radiated away from the red hot decarbonating and clinkering zones directly into the colder parts of the kiln without doing any useful work. In addition to this, much high grade heat is carried away bodily by the hot gases from the decarbonating zone passing into the colder zones

Now it is shown that every 918.6 B Th U 's of high grade heat thus passing away into the colder parts of the kiln represents a loss of 1 lb of potential clinker

It is proved that if in an ordinary rotary kiln in which we are burning 30 tons of coal per 100 tons of clinker made, we suddenly stopped all internal radiation, our output of clinker would rise from 100 to 273 tons and if we next stopped all external radiation as well from the kiln, the output of clinker would rise to $273 + 15 = 288$ tons of clinker. It is assumed here that the entering air is preheated to 400°F (which is the normal temperature of preheating in a rotary kiln)

Summary

| | Coal Burnt | Clinker Made |
|--|------------|--------------|
| The ordinary rotary kiln | 30 tons | 100 tons |
| After stopping internal radiation | 30 tons | 273 tons |
| After stopping internal and external radiation | 30 tons | 288 tons |

Method of Preventing Internal Radiation Losses

It will be seen therefore that enormous losses occur inside a rotary kiln owing to internal radiative losses. The question is how are these losses to be stopped?

In order to attempt the solution of this question it is essential to understand the causes of the great internal radiation losses of the rotary kiln

Undoubtedly the main cause is the fact that in the rotary kiln there occurs a great volume of empty space, or rather, space filled only by thermally transparent gases. Consequently, high-grade heat can freely radiate through these hollow spaces from the hotter to the colder parts of the kiln.

We are helped considerably in seeing how this inherent defect of the rotary kiln may be overcome by considering the case of the shaft kiln. Here internal radiation and convection losses are largely reduced by the device of filling the interior with a mass of coke and raw-material briquettes, so that any radiant or convective heat is caused to impinge on a surface, and consequently high-grade heat cannot pass from hotter to colder places without doing useful work by heating intervening raw material. This fact is at once reflected by the greater thermal efficiency of the shaft kiln. Whereas, by burning 30 tons of coal in the rotary kiln we only obtain, say, 100 tons of cement; in the shaft kiln for the same amount of fuel we obtain from 150 to 190 tons of cement clinker. Hence, by introducing baffling in the shaft kiln, we have increased the clinker output by 50 to 90 per cent. for the same amount of heat expended.

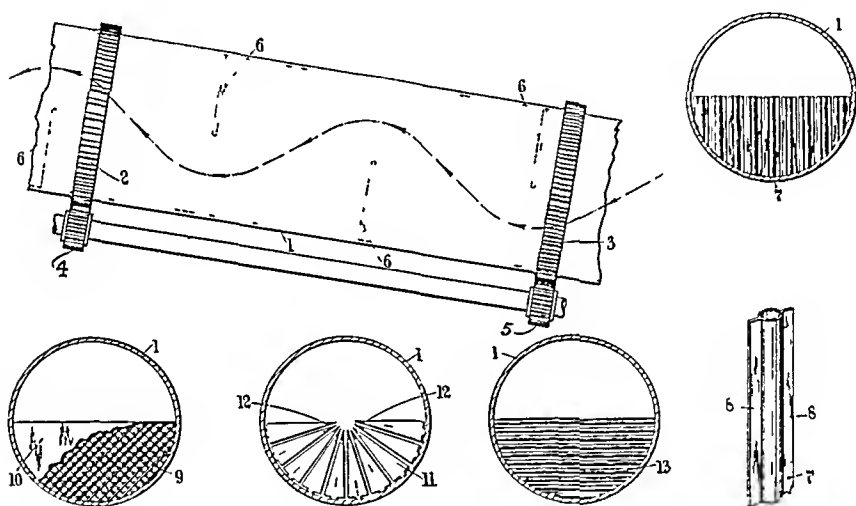


FIG. 2.—Increasing Efficiency of Rotary Kiln with Introduction of Baffles.

Increasing the Efficiency of the Rotary Kiln by Baffling

Now one method of increasing the efficiency of the rotary kiln is by introducing baffles in the manner described in Patent (British Patent 264,920 (application date, 27th October 1925; accepted, 27th January 1929)) by the writer and J. Darnley Taylor.

The idea consisted in placing inside the kiln a considerable number of baffles, so arranged as to shield the different parts of the kiln from direct radiation losses.

Fig. 2 shows a rotary kiln thus fitted with baffles, these being placed in staggered relationship to each other, whereby the kiln is divided into a plurality of chambers, and loss of heat by direct radiation from one part of the kiln to another is largely prevented. Also, since in general the gases flowing through a kiln do not intermingle to any great extent with the raw material but escape at a high temperature, this arrangement is also beneficial in facilitating heating by mutual physical contact—of gas and material—the main method of heating at low tem-

peratures but not at high temperatures where the main heating is effected by radiation from hot surfaces

However by adopting these devices we are brought up against the serious practical difficulties. The valuable effective B T H U s lost by radiation are precisely those evolved at a very high temperature so that the baffling media must be able to withstand not only a white heat but also the heavy pounding action of the clinker passing down the kiln.

Then again the introduction of baffles spoils the draught of a kiln and makes necessary the introduction of fans etc. at the cold end preferably attached to waste heat boilers.

Undoubtedly great improvements are possible in this direction but the technical difficulties to be overcome are considerable.

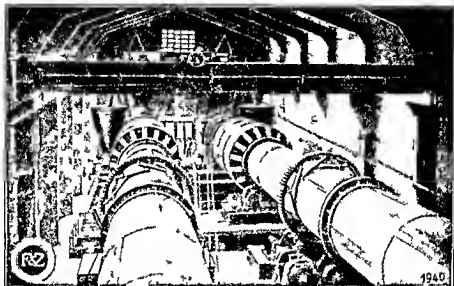


FIG. 3 A Dry process Kiln with Enlarged Calcination Zone

Increasing the Efficiency of Rotary Kilns by Means of Enlarged Calcination Zones *

We have seen in the preceding pages that the main thermal losses occur by internal radiation losses and by convection losses principally from high grade heat. We have seen that there corresponds a loss of 1 lb. of clinker to every 918.6 B T H U s available above 1481°F (805°C) which is lost inside the kiln in the decarbonating and sintering zones. Now one way of diminishing the losses by internal convection of the hot gases is to make the gases traverse the decarbonating zones of the kiln as slowly as possible without however reducing the output from a given kiln.

One way of achieving this is to *enlarge* the calcination zone. This not only reduces the convection losses (by allowing the hot gases time to communicate their heat to the raw material) but also tends to reduce internal radiation losses as well.

* Reproduced from *Cement and Cement Manufacture* December 1925 by the courtesy of the Editor.

For several years past the specially constructed type of rotary kiln described below has been in successful operation on the Continent. The rotary kilns referred to are fitted with enlarged decarbonating zones—which must not be confused with the enlarged sintering zone of ordinary kilns.

The zone built into the kiln under discussion is of a very much greater diameter than that of the normal cement kiln, and in this enlarged zone the material is not “sintered” but only “decarbonated,” *i.e.*, heated to the correct temperature to expel the carbon dioxide from the mass, but not to a temperature high enough to cause the formation of cement clinker. It is a well-known fact that the raw material in a kiln—especially in the dry process—shows a tendency to “rush” forward in an irregular manner, with the consequence that the output of the kiln is seriously reduced and the resulting clinker shows irregularity of quality. It is obvious that the raw material traversing a kiln in the ordinary way can never be uniformly treated so that every particle is in exactly the right condition to enter the sintering zone and be at once converted into cement clinker. This is due to the fact that in the dry process the material tends to traverse the kiln in a series of “rushes,” so that material often reaches the sintering zone before it is properly decarbonated, whilst in the wet process there are not only small particles but also medium-sized particles, and even quite large masses of raw material from slurry rings and clinker rings, impinging into the sintering zone. The larger particles travel through the kiln at speeds different from those of the smaller particles, with the result that the particles arrive in the sintering zone unequally calcined.

All these deficiencies are claimed to be overcome by the enlarged calcining zone built into the kiln just before the sintering zone. The raw material collects in considerable quantities in this zone, and its progress through the kiln is arrested for an additional two hours. The consequence of this is that the material is uniformly and thoroughly calcined, ready and properly prepared to go into the sintering zone. It is therefore easy to see that by introducing such well-calcined materials into the sintering zone much larger quantities can be sintered in a given time, and as a result the coal consumption will be much reduced. The following gives some recent data on this point:—

Dry Process.—Two kilns were erected at the Königshofer cement factory in Königshof. Their sizes were 3 metres (9 ft. 10 in.) diameter by 50 metres (164 ft.) long. Each was fitted with an enlarged calcining zone 5 metres (16 ft. 5 in.) diameter by 9 metres (29 ft. 6 in.) long. Both kilns operated on the dry process under forced draught. The exit gases escaped at a temperature of 650° to 700° C. (1,200° to 1,300° F.) and were passed through a waste-heat boiler. These waste gases developed 4,000 to 4,400 kw. (5,000 to 5,900 H.P.), *i.e.*, more power was produced than sufficed to run the whole works, and part of this power was delivered to another works. The output of clinker was 280 to 300 tons per twenty-four hours, in spite of the fact that the raw material was very high in calcium carbonate (78 per cent.). The coal consumed varied between 1,350 and 1,400 calories per 1 kg. clinker (19 to 20 tons of standard coal per 100 tons of clinker).

Wet Process.—One kiln of 3 metres (9 ft. 10 in.) diameter and 50 metres (164 ft.) long, fitted with a calcining zone 5 metres (16 ft. 5 in.) diameter by 9 metres (29 ft. 6 in.), gave an output of 270 to 280 tons in twenty-four hours, with a coal consumption of 1,600 to 1,700 calories per kilogram of clinker (23 to 24 tons of standard coal per 100 tons of clinker). It should be mentioned that the raw materials treated in this kiln were admittedly the most difficult materials in Germany to burn.

In another place, where better raw materials were at hand, the kiln yielded

300 tons and more with a coal consumption of about 1 600 calories per kilogram clinker (23 tons of standard coal per 100 tons of clinker)

Several ordinary rotary kilns of the same dimensions, viz , 3 metres diameter (9 ft 10 in) by 50 metres (164 ft) which were working side by side with the above mentioned two kilns fitted with the enlarged calcining zones and using exactly the same raw materials had an output of 210 tons per twenty four hours, with a consumption of 33 to 34 tons of standard coal per 100 tons of clinker, the coal having a calorific value of 7 000 calories (12 600 B Th U s per lb), that is, about 2,300 to 2 400 calories per kilogram of clinker It will be seen from this difference in fuel consumption and output that the enlarged calcining zone was here very successful

Other kilns now operating give similar outputs and coal consumptions It has been proved that on the average the introduction of these enlarged calcining

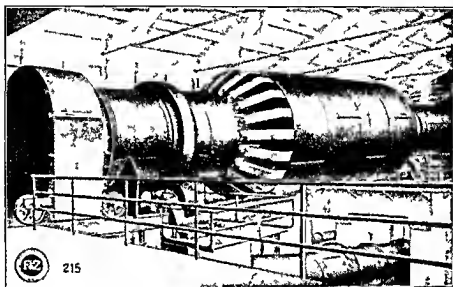


FIG 4 —Dry process Kiln at Königshof with Enlarged Clinker ng Zone

zones leads to an increase of output of about 30 per cent and a diminution of coal consumption of about 25 to 30 per cent Moreover it is claimed that in the kilns fitted with the enlarged calcining zone a far better burnt clinker is obtained than has hitherto been possible while crushing and tensile strengths are increased by 20 to 30 per cent due to the fact that the raw material, which must pass through the decarbonating zone is held up in this zone for two hours longer than usual and thereby comes into the sintering zone far more completely decarbonated than is the case with the ordinary rotary kiln

Baffling Radiant Heat by Means of Dust

Meanwhile attempts to solve the problem along other lines were being made

Instead of using solid diaphragms to act as baffles at a white heat why not use the raw material itself in the form of dust ? This would stand up to the clinkering temperature itself, and it is well known that finely divided dust presents an enormous surface for arresting radiation For example, accurate experiments

carried out by the writer have shown that 1 lb. of quartz sand finely ground in a tube mill presents a surface running from 300 to 2,000 sq. ft.*

It is well known what an effective arrester of radiant energy is formed by a dust cloud or fog. A certain weight of water dispersed in the form of very fine globules will arrest light or radiant energy far more effectively than if this water is condensed into a single transparent mass.

Consequently, filling the interior of a kiln with a mass of raw material reduced to a state of extremely fine division will stop the internal radiation losses going on in the ordinary rotary kiln 'as effectively as is now achieved in the shaft kiln, by filling up its interior with large blocks of raw material and coke.

Davis's Experiments

Among the first to try practical experiments on a large scale on these lines was Arthur Charles Davis of the British and Associated Portland Cement Co. His apparatus (British patent specification No. 307,763; application date, 11th October 1927; complete accepted, 11th March 1929) is shown in Fig. 5.

The apparatus consisted of a vertical-shaft kiln at the base of which slurry or raw meal is injected as a fine suspension through one or more jets located at or adjacent to the base of the kiln, the injection being vertically up into the kiln through the plane of the combustion zone, so that the material has a passage upwards and downwards through the hot gases.

The velocity at which the slurry is projected into the kiln and its degree of atomisation or subdivision may be so arranged that the slurry or other material is kept in suspension by the velocity of the air until its density is increased by calcination to such an extent as to cause the grains to fall to the base of the kiln. The flame temperature could also be adjusted so that in combination with the degree of atomisation of the slurry the period occupied in falling through the flame is sufficient to raise the temperature of the material to the clinkering temperature, which is in the neighbourhood of 1,400° C. (2,552° F.). That is to say, with coarsely atomised slurry the period for transmission of the heat throughout the particles will be longer, and in such a case the flame temperature must be raised, and vice versa.

Also, the flame may be adjustable, so that if necessary a swirling or rotary

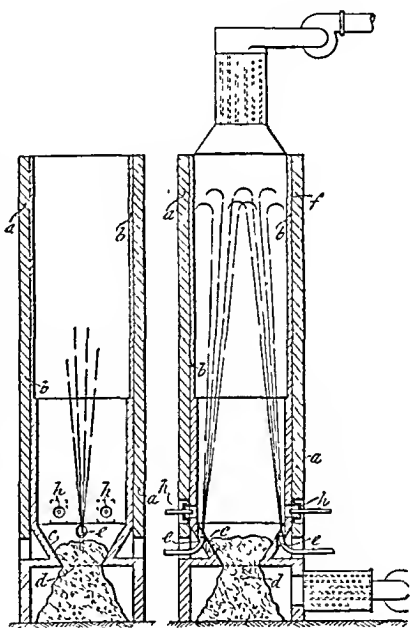


FIG. 5.—Filling Interior of Kiln with Raw Material to Reduce Radiation Losses.

* See the following papers by the writer and his co-workers: "Researches on the Theory of Fine Grinding," Part III., *Trans. Ceramic Society*, 1925-26, vol. xxv., Test Nos. 145-151. "Connection between the Surface Area Produced and the Work Done in Tube-mill Grinding of Quartz Sand." By Geoffrey Martin, Edgar A. Bowes, and F. B. Turner. "Researches on the Theory of Fine Grinding," Part X.; "On the Connection between the Statistical Diameter of Crushed-sand Particles and their Statistical Surface," *Trans. Ceramic Society*, 1927-28, vol. xxvii., pp. 59 *et seq.* By Geoffrey Martin and Edgar A. Bowes.

countercurrent manner, the coldest raw material thus coming into contact with the coldest issuing gas, and the gas being finally passed through a waste-heat boiler. Fig. 6 shows Girouard's installation. The vertical kiln shaft A is sunk with its longitudinal axis approximately parallel to the face B of a quarry, the base of the shaft proper being on the ground level B² of the quarry.

A working adit B communicated with the base of the shaft and thus gives access to the burners C, to which pulverised fuel is supplied through one or more pipes D.

The raw materials, after crushing, drying, and storing in a series of bins F (which may be excavated in the ground), are transferred by means of conveyors and elevators to a hopper or bin J, of a pulverising apparatus J, whereby the materials are introduced into the top of the kiln in the form of a dense cloud of dry pulverised particles.

Around or between the burner nozzles C are by-passes A² through which the hot clinker passes into a cooling shaft K sunk in the ground below the kiln shaft and working adit, whence by means of a series of conveyors and elevators M the formed clinker can be removed. The burners C are supplied with hot air by means of a pipe L, which passes through the hot clinker.

Q is a waste-heat boiler, P a dust collector, and R¹ a fan.

Technical Difficulties due to Escape of Dust—Martin's Researches

Although this apparatus represents a considerable advance on Davis's kiln, there still remained the serious technical difficulties to be overcome of preventing the dust being blown out of the kiln by the upward rush of hot gas.

There is a definite connection between the velocity of the gas and the size of the particle which it will carry out. The connection was worked out by the writer.*

It must be recollected that it is impossible to generate cement clinker without generating gas at the same time.

For example, in the *wet process* of cement formation, to every 1 lb. of clinker produced no less than 5 lbs. of gas are generated, which occupy 167 cub. ft. of space at an exit temperature of 800° F., while in the *dry process* about half this weight of gas is generated, which at an exit temperature of 1,350° F. occupies about 100 cub. ft.

It is obvious, therefore, that if a large output of clinker is aimed at, there will be a vast volume of gas rushing out of the kiln at a great speed. In the ordinary rotary kiln the gas leaves the kiln at a speed of about 16 ft. per second.

Such an outrush of gas will carry with it great volumes of dust, and, indeed, prevent the entry of fine dust into the kiln at all, as it will be blown out faster than it enters. It also renders the output of the kiln small. The particles must descend against a stream of ascending hot gas, and consequently the particles must be of the correct size. A great deal of research work was carried out by the writer with a view to deciding the correct weight and size of the particles to introduce. If the particles are too small, they will float away as smoke, and if too large, they will fall too rapidly down the kiln and probably emerge as underburnt clinker. The correct limits of size, with due allowance for viscosity of the hot air, were worked out, and the writer succeeded in completely overcoming these difficulties by introducing a cyclonic action in the upper part of the kiln as described in British Patent No. 276,066 (application date, 17th May 1926; complete acceptance, 17th August 1927).

* See the following papers: "Researches on the Theory of Fine Grinding, Part VI.: On the Diameters of Irregularly Shaped Crushed-sand Particles Lifted by Air Currents of Different Speeds and Different Temperatures," *Trans. Ceramic Society*, 1926-27, vol. xxvi., p. 21. By Geoffrey Martin. "Researches on the Laws of Air Elutriation," *Trans. of the Institution of Chemical Engineers*, 1926, vol. iv., p. 164. By Geoffrey Martin.

The Geoffrey Martin Flotation Kiln

The following is a description of the kiln (The kiln was sunk in the ground) The cyclonic action was induced in the upper part of the kiln by making it considerably wider than the lower clinkering zone, and so arranged that the hot gas from the clinkering zone impinges at a high speed upon the mass of gas in the upper zone tangentially along one side, as indicated in Fig 7 below

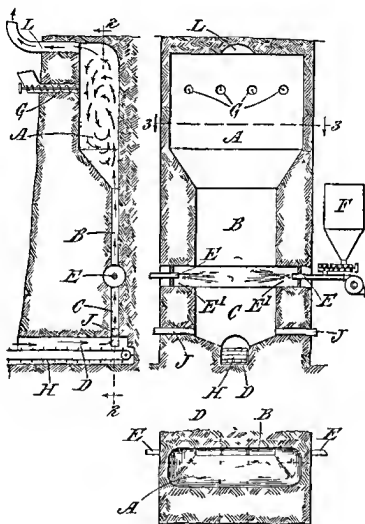


FIG 7 —Geoffrey Martin Flotation Kiln

The effect of this arrangement is that the stream of hot gas rushing up the clinkering zone impinges eccentrically upon a slower moving stream of colder gas in the calcining zone with the result that a rotary movement is set up in the larger volume of gas in the calcining zone, which results in a downward movement of gas on the colder side of the kiln remote from the clinkering zone and an upward movement of gas on the hotter side of the kiln closer to the clinkering zone

Also the introduction of pulverised raw material is arranged to take place in such a manner that it is in the *same* direction as the downward gaseous current

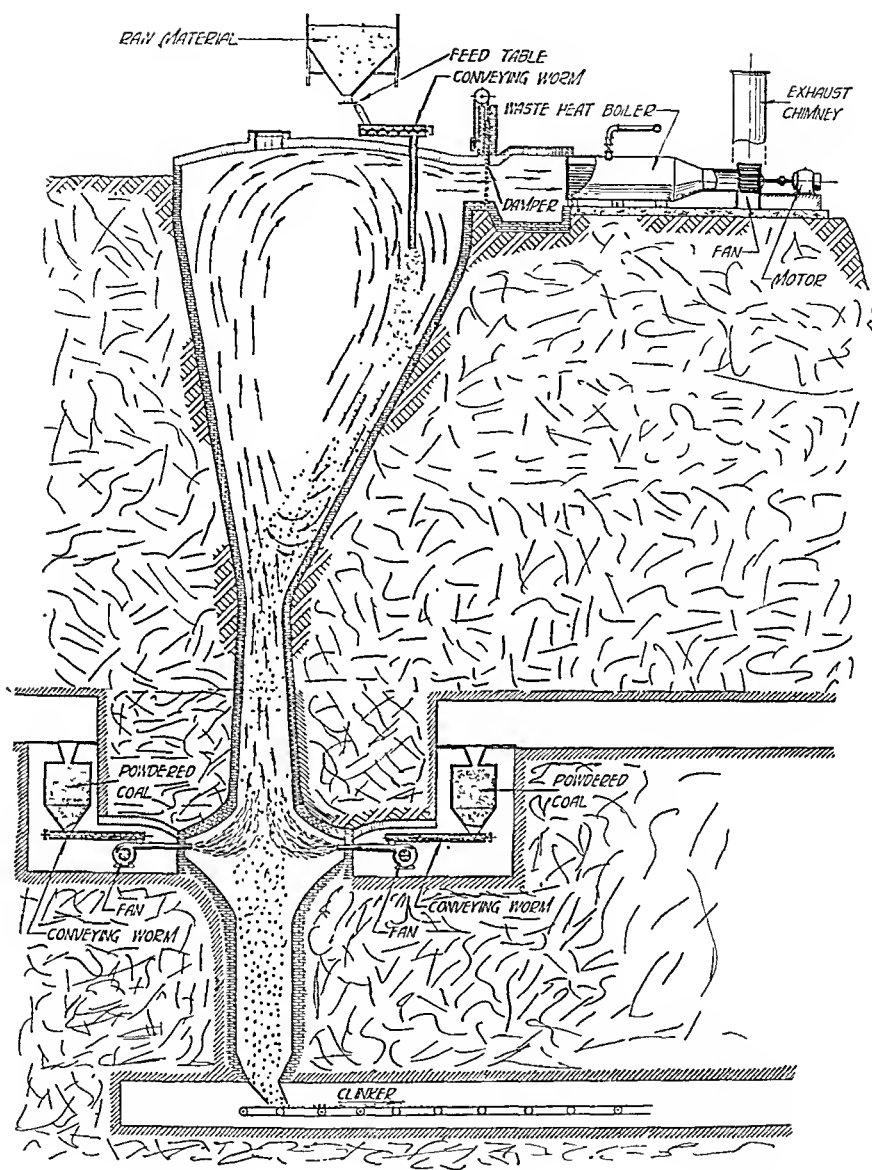


FIG. 8.—Diagram of Kiln Described by Dr Martin.

This is achieved by introducing the cold raw material at the side of the shaft—as shown in the figure. This causes the kiln to be cooler on one side than on the other, with the result that a circulatory and downward movement of the hot gas takes place from the hotter side of the kiln to the colder side, which sweeps the finer particles downwards towards the hotter parts of the kiln, where, becoming

incandescent, they sinter or fuse together into larger particles and thus are not swept out of the kiln

In order to accentuate still further this circulatory and downward movement of gas at the point of entry of the raw material into the shaft, the exit gas is drawn off from a point on the same side as the side on which the raw material enters, but above it

The exit gases are passed through a waste-heat boiler by suction from a fan

In effect, by this arrangement the upper part of the kiln is made to act like a gigantic cyclonic dust collector, only a negligible amount of microscopically fine dust escaping through the waste heat boiler into the chimney

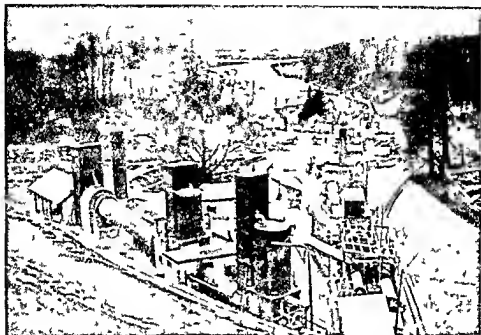


FIG. 9—Coal pulverising Plant and Raw material Dryers at Asheham

Problem of Obtaining a Large Output from a Dust or Flotation Kiln

In the early part of this article it was pointed out that a cement kiln, in order to be commercially successful, must (1) be thermally efficient, and (2) must possess a large output. Of the two conditions probably the second is the more important commercially

It has also been pointed out that the output of clinker from a cement kiln is proportional to the volume of gas pouring through it, as the production of each lb. of clinker is necessarily attended with the production of a definite volume of gas. Any furnace in which the gas cannot freely flow through will necessarily have a small output and vice versa

For example, gas is impeded from passing through an ordinary shaft kiln because its interior is filled with solid blocks, or briquettes, and the combustion gas has to pass between their interstices, thus exposing the gases to great friction in their tortuous passage

Therefore, the output from a shaft kiln is low, for only a small amount of gas

can traverse it in a given time. On the other hand, gas can freely pour through a rotary kiln, and this is why its output is so large.

Now in the case of a kiln of the present type gas can freely traverse it because of the smallness and consequent mobility of the particles filling it—the dimensions of the particles being carefully confined so as to cause them to fall in the ascending flue gas at the same speed as they are uplifted, *i.e.*, the terminal speed or fall of the particles is made equal to the speed of the gases traversing the kiln. The particles thus *float* in the ascending gases and do not impede the flow of the latter through the kiln. Consequently, a great volume of hot gas may be made to traverse the flotation kiln in a given time, and, therefore, as great an output of clinker

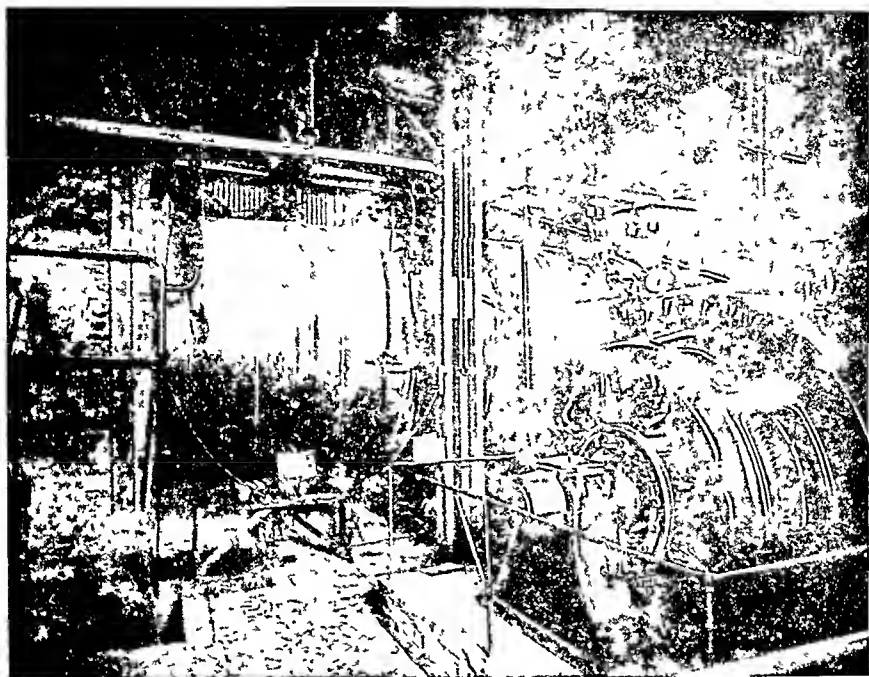


FIG 10 —Waste-Heat Boiler at Top of K.ln.

is attainable in the flotation kiln as in the rotary kiln, and for the same reason, the ease with which hot gases can traverse both types of kiln.

In fact, this flotation kiln may be regarded as a particular case of a shaft kiln in which the preheating and calcining zones are filled, not with the usual large blocks of material to be burnt mixed with fuel, but with small particles of raw material of such a size that they are easily lifted by an uprising gaseous velocity of 4 ft. per second, but sink in a gaseous velocity of 1 in or less per second, so that the contents of the calcining zone form a coarse powder which is maintained in a state of agitation and is heated by hot gases of combustion traversing it; and this powder does not materially impede the progress of the gases through it, since the particles are being continually swept out of the way by the ascending volumes of gas. Hence, as large volumes of gases can pass in a given time, a large output of clinker in a given time is conditioned by the volume of gas which can escape in this given unit of time.

Improved Mixing of Raw Materials in Flotation Kiln

As is well known, a frequent cause of unsoundness in finished cement is inefficient mixing of the raw materials before entering the furnace. In fact this is the main reason why cement manufacturers prefer to add water to the raw materials in order to make them into a thin mud (in spite of the fact that all this water must be later evaporated again and consume fuel thereby). Now in the ordinary wet process of cement manufacture this mixing is achieved by grinding the materials when suspended in the fluid water, since mixing when suspended in a fluid is always much more perfect than when the materials are mixed dry.

In the flotation kiln process this mixing is effected by mixing the powder in the fluid gases of the kiln itself in the upper cyclone portion of the kiln, whereby a much more perfect mixture is achieved, and when the mixing is completed

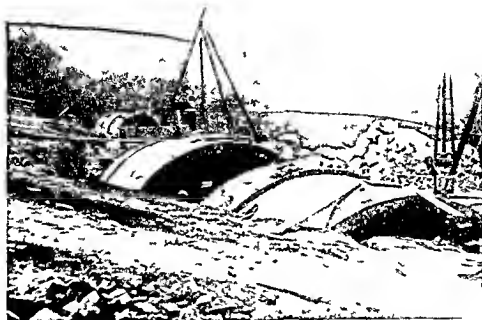


FIG. 11.—Top of the Furnace in Course of Erect on during Asheham Experiment

the gases pass away themselves without the necessity of burning extra fuel to distil them away, as occurs when water is used as the mixing fluid.

The mixing occurs in the decarbonating and preheating part of the kiln the particles being whirled around by the circulating currents of hot gas in these zones until complete admixture is effected before they can pass into the clinkering zones. This process of mixing without the aid of water by using a thin fluid like hot gas represents a considerable technical advance.

Other Advantages of the Flotation Kiln

A continuous process where possible, is almost always much more economical than a discontinuous process especially as regards furnaces, where masses of brickwork must be heated.

In this respect the flotation process kiln is like the rotary kiln, quite continuous

in operation. The raw material in the form of a powder is fed in at the top of the kiln, and continuously issues as clinker at the bottom end. The process goes on continuously night and day, so long as the supply of raw material is kept up and the fuel is kept burning in the burners.

The flotation kiln possesses advantage over the rotary kiln in that there are no moving parts to get out of order. In a rotary kiln a great mass of iron and brickwork, weighing many hundreds of tons, must be kept moving continuously, supported on massive steel rollers. In the flotation kiln all these disadvantages disappear because the kiln itself is stationary, and it is simply the action of gravity which pulls the material down and through the kiln. Moreover, in the rotary kiln the rotation of the brickwork causes the formation of clinker rings, which occasion periodical losses and stoppages. In the flotation kiln all such movements are absent as would cause the formation of clinker rings, so that such a kiln may be expected to run continuously without any interruptions for months at a time. The action of the linings also must, for similar reasons, be reduced to a minimum.

Erection of a Large Flotation Kiln at Asheham, near Lewes, Sussex, England

In order to test out the advantages of a flotation kiln over a rotary kiln, and the possibility of manufacturing clinker on a large scale, a large-scale experiment was carried out at Asheham by a group of financiers, among whom may be mentioned Charles F. Lumb, Sir Percy Girouard, Mr Hamilton, and S. R. Worley, all of London.

The kiln was built by the writer, being about 150 ft. deep and possessing an area of about 800 sq. ft. at the top. It was fitted with a waste-heat boiler at the top.

A diagrammatic view of the kiln is shown in Fig. 7. Photographs of the works are also shown.

A series of trials carried out during December 1929 showed most promising results. It is hoped, after making certain alterations, to remedy technical defects which revealed themselves in the trials, to proceed to manufacture clinker by this new process in the near future.

Lellep's Process

Another method of attacking the problem is due to Lellep:—

1. The dry and finely ground material is made into small balls by granulating in a rotating drum in the presence of drops of water. The drops of water cover themselves with meal, and, as the drum rotates, increase in size to about $\frac{1}{2}$ in. diameter.

2. The granules are now fed from a storage hopper on to a travelling grate, where they are calcinated and the carbon dioxide partially expelled therefrom without the material being sintered.

The heating is effected by the exit gases from the rotary kiln used in the final stage to sinter the material. The hot gases meet the material on the grate above, and pass down through it, being sucked through the grate. The apertures between the links of the grate are so small that no granules pass through.

The upper layers of meal on the grate are heated by the exit gases to such an extent that they are completely decarbonated. The lower layers, however, still retain carbon dioxide, and serve to protect the grate from burning away, because the gases are cooled in the process of decarbonating the upper layers. The average temperature of the gases, after having passed through the grate, is only just over 100° C. (212° F.).

It is only near the discharge end of the grate that the temperature of the gases

actually coming into contact with the grate reaches a temperature as high as 550°C ($1,022^{\circ}\text{F}$) So that burning of the bars and special cooling of the grate is unnecessary

3 The partially decarbonated material now falls down a shoot into a rotary kiln fired in the ordinary way with powdered coal It is in this rotary kiln that the final sintering is effected (since proper sintering on a travelling grate is very difficult to achieve) The rotary kiln is only about one third of the usual length, since it merely completes the burning and sintering of the raw material In fact, the rotary kiln is only slightly longer than the ordinary burning zone The sintered material passes from the kiln through a cooler in the ordinary manner

4 The dust carried up the rotary kiln by the gases is caught in the layer of granules on the travelling grate, which thus acts as a filter The gases entering the chimney are thus practically dust free

5 The clinker made was equal to the best rotary cement clinker

6 In one experimental plant at the Ruedersdorf Portland Cement Plant, using coal of 6,450 calories per gram (11,600 B Th U's per lb), raw meal of 76.5 per cent CaCO_3 and of moisture content 13.5 per cent, it was found that the heat consumed averaged 1,045 large calories per kilogram of clinker made, i.e., 100 tons of clinker could be produced by the consumption of about 15 tons of standard coal of 12,600 B Th U's per lb Thus the fuel consumption was much less than that experienced with rotary kilns (28 to 33 per cent of coal), and even lower than that of vertical shaft kilns (say, 18 per cent)

In this connection (as with all dry process kilns) the initial drying of the raw material is costly and difficult, and the fuel consumption in the preliminary drying of the raw material, previous to making the meal should be reckoned on to the fuel costs of the process

The plant is placed on the market by G. Polysius, A.G., Dessau

A fuller description than is here possible is given in *Cement and Cement Manufacture* of May 1931, p. 532, in an article entitled 'The Burning of Cement on a Travelling Grate'

It will be obvious from the contents of this chapter that attempts to improve the manufacture of cement are still actively in progress

CHAPTER XXIX

SOME USEFUL DATA AND PRACTICAL RULES FOR
CEMENT ENGINEERS

CHAPTER XXIX

Useful Data for Cement Engineers

1 lb. standard coal yields 12,600 B.Th.U.'s=7,000 kilo-calories.

12,600 B.Th.U.'s per lb. is the same as 7,000 calories per gram.

1 ton standard coal yields 7.112×10^6 kilo-calories.

1 American Barrel=380 lbs. (old style)=376 lbs. (new type).

1 German " Fass "=170 kilos of cement=374.75 lbs.

To convert quantities relating to American " Barrels " to those relating to 100 tons (long=2,240 lbs.), multiply by the factor 5,895.

To convert kilo-calories per 100 kilograms of clinker to B.Th.U.'s per 100 tons clinker, multiply the kilo-calories by 4032.3.

To convert kilo-calories per kilogram to B.Th.U.'s per lb., multiply the number of kilo-calories by 1.8.

To convert kilo-calories per cubic metre to B.Th.U.'s per cubic foot, multiply the kilo-calories by 0.11236.

To convert x lbs. weight per cubic foot to kilograms per cubic metre, multiply the x lbs. by 16.0184.

Rough Engineering Rules for Cement Engineers

Ordinary boiler with ordinary furnace—

1 sq. ft. *heating surface*=7 to 8 lbs. steam/hour.

Waste-Heat boiler (wet process)—

1 sq. ft. *heating surface*=1.5 lbs. steam/hour.

1 ton clinker/hour= $\frac{1}{2}$ ton steam/hour.

Waste-Heat boiler (dry process)—

1 sq. ft. *heating surface*=3 lbs. steam/hour.

1 ton clinker/hour=1 ton steam/hour.

Requirements of power, ordinary kiln—

1 ton coal (all purposes)=2 tons cement.

1 ton cement/hour=1 ton steam/hour necessary.

Output per week in clinker *tons*= $1.8 \times H.P.$ *required*.

Table of Fuel Consumptions of Cement Kilns Expressed in Various Units

Calorific value per lb. expressed in calories = 1.8 X calorific values in B Th U 's

1 kilogram = 2.2046 lbs

1 kilo calorie = 3.9683 B Th U 's

1 B Th U = 0.252 kilo calorie

1 lb. standard coal = 12,600 B Th U 's when burnt

1 kilogram standard coal = 7,000 kilo calories when burnt

| Cwts of Standard Coal per 1 Ton Clinker | Tons of Standard Coal per 100 Tons Clinker | B Th U 's per 1 Lb Clinker | Kilo calories per 1 Kilo Clinker |
|--|--|-------------------------------|--|
| 30 | 15 | 1890 | 1050 |
| 32 | 16 | 2016 | 1120 |
| 34 | 17 | 2142 | 1190 |
| 36 | 18 | 2268 | 1260 |
| 38 | 19 | 2394 | 1330 |
| 40 | 20 | 2520 | 1400 |
| 42 | 21 | 2646 | 1470 |
| 44 | 22 | 2772 | 1540 |
| 46 | 23 | 2898 | 1610 |
| 48 | 24 | 3024 | 1680 |
| 50 | 25 | 3150 | 1750 |
| 52 | 26 | 3276 | 1820 |
| 54 | 27 | 3402 | 1890 |
| 56 | 28 | 3528 | 1960 |
| 58 | 29 | 3654 | 2030 |
| 60 | 30 | 3780 | 2100 |
| 62 | 31 | 3906 | 2170 |
| 64 | 32 | 4032 | 2240 |
| 66 | 33 | 4158 | 2310 |
| 68 | 34 | 4284 | 2380 |
| 70 | 35 | 4410 | 2450 |
| 72 | 36 | 4536 | 2520 |
| 74 | 37 | 4662 | 2590 |
| 76 | 38 | 4788 | 2660 |
| 78 | 39 | 4914 | 2730 |
| 80 | 40 | 5040 | 2800 |
| 82 | 41 | 5166 | 2870 |
| 84 | 42 | 5292 | 2940 |
| 86 | 43 | 5418 | 3010 |
| 88 | 44 | 5544 | 3080 |
| 90 | 45 | 5670 | 3150 |
| 92 | 46 | 5796 | 3220 |
| 94 | 47 | 5922 | 3290 |
| 96 | 48 | 6048 | 3360 |
| 98 | 49 | 6174 | 3430 |
| 100 | 50 | 6300 | 3500 |
| 102 | 51 | 6426 | 3570 |
| 104 | 52 | 6552 | 3640 |
| 106 | 53 | 6678 | 3710 |
| 108 | 54 | 6805 | 3780 |
| 110 | 55 | 6930 | 3850 |

Coal

1 lb. of standard coal (12,600 B.Th.U.'s per lb.) = $\frac{100}{100-y} \times \frac{12600}{x}$ lbs. of commercial coal bought as developing x B.Th.U.'s per lb. and containing y per cent. moisture or loss.

To Express Ordinary Moist Coal as Bought in Terms of Standard Coal

Step 1.—Calculate the *standard coal factor* (S.C.F.) of the coal as bought by the formula—

$$\text{S.C.F.} = \frac{100-y}{100} \times \frac{x}{12600},$$

where x = B.Th.U. value of the coal in question (determined dry);
 y = the moisture of the coal as bought.

Step 2.—Multiply the greater number of tons of the moist coal as bought by the S.C.F.

Example.—100 tons of cement are made from 50 tons of coal “as bought” containing 10 per cent. of moisture and of B.Th.U. value 10,000. Reduce this to standard coal.

Here,

$$\text{Per cent. moisture} = y = 10,$$

$$\text{and B.Th.U. value of coal} = x = 10,000.$$

$$\therefore \text{S.C.F.} = \frac{100-10}{100} \times \frac{10000}{12600} = 0.7142.$$

Hence tons of standard coal consumed per 100 tons clinker made are $50 \times 0.7142 = 35.7$.

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Directions for Using Index

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